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Demonstration of Wellman-Lord/Allied Chemical FGD Technology: Demonstration Test Second Year Results

TRW, Inc.
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DEMONSTRATION OF WELLMAN-LORD/ALLIED CHEMICAL FGD TECHNOLOGY

Final Report and Demonstration Test Second Year Results

by

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16. ABSTRACT The report gives results of an evaluation of the performance (over a 2-year period) of a full-scale flue gas desulfurization (FGD) unit to demonstrate the Wellman-Lord/Allied Chemical process. The process is regenerable, employing sodium sulfite wet scrubbing, thermal regeneration of the spent scrubber solution, and reduction to elemental sulfur of the recovered SO2. Tests indicate that 89-90% of the SO2 can be removed readily from the flue gas in a long-term dependable manner. However, best performance of the regeneration system for 1 year was limited to only 61% of the time called upon to operate. Major operating limitations were reduction unit problems, but unscheduled evaporator and booster blower outages, as well as start-ups and shutdowns, also contributed to downtime. As expected, process energy requirements, primarily for thermal regeneration of the scrubber solution and subsequent recovery of SO2, were quite large, amounting to 12% of the boiler heat input derived from fuel. Actual annualized operating cost was 14.9 mills/kWh, at 1978 prices for raw materials and utilities. Credits for the sale of byproduct sulfur were only 0.2 mills/kWh. The reported operation and performance occurred after some modification to the boiler to increase inlet flue gas temperature and after implementing FGD plant improvements identified during initial operation.

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Flue Gases	Coal	Stationary Sources	21B	21D
Desulfurization		Wellman-Lord/Allied Chemical Process	07A, 07D	
Scrubbers			13I	
Sodium Sulfites			07B	
Sulfur				

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ABSTRACT

Performance of a full-scale flue gas desulfurization unit to demonstrate the Wellman-Lord/Allied Chemical process was evaluated for a period in excess of two years. The Wellman-Lord/Allied Chemical process is a regenerable process employing sodium sulfite wet scrubbing, thermal regeneration of the spent scrubber solution, and reduction to elemental sulfur of the recovered sulfur dioxide.

Test program results indicate that 89 to 90% of the SO_2 can be readily removed from the flue gas in a long-term, dependable manner. Reliability of the coupled absorber and regeneration system for the second year was 61%. For the last seven months, it was 74%. The major limitations to operations were reduction unit problems but unscheduled outages of the evaporator and the booster blower and startups and shutdowns also contributed to down time.

As expected, the energy requirements of the process, primarily for thermal regeneration of the scrubber solution and subsequent recovery of SO_2, were quite large and amounted to 12% of the boiler heat input derived from fuel. Actual annualized operating cost was 14.9 mills/kWh, using 1978 prices for raw materials and utilities. Credits for the sale of byproduct sulfur amounted to only 0.2 mills/kWh.

The reported operation and performance occurred after some modification to the boiler to increase inlet flue gas temperature and after needed improvements to the FGD plant identified during initial operation were implemented. Design limitations affecting overall performance were lack of redundancy, regeneration area capacity of only 80% of full load, underdesign of the purge solids dryer, and limited turndown capability.

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CONTENTS

Page

Abstract. ii

Figures vii

Tables. x

Acknowledgement xv

Executive Summary xvi

1. Introduction. 1

  Background 1

  Program Goals. 2

2. Program Overview. 4

  Chronology of Events 4

  Summary of Results 8

    Demonstration Testing 8

    Baseline Testing. 11

    Acceptance Testing. 13

  Test Planning. 15

  The Two-Year Demonstration 20

    Process Description 24

    Performance Evaluation Methodology. 28

TABLE OF CONTENTS (continued)

Page

3. Discussion of Results 32

Sulfur Dioxide Removal 32

Removal Performance 35

Feed Characteristics. 40

Flue Gas Flow Rate. 45

Inlet Concentration of SO_2. 48

FGD Plant Dependability. 50

Reliability 54

Operability 56

Utilization 60

Reasons for Interruptions. 60

Design Limitations 64

Minimum Sustainable Load. 65

Absorber Turndown 68

Maximum Limitations 68

Load Following Capability 71

Purge Treatment Limitations 73

Energy Consumption 81

Steam 82

Electrical Energy 87

Natural Gas 88

Process Economics. 89

Raw Material Consumption. 94

Byproduct Production. 95

TABLE OF CONTENTS (continued)

Page

Pollutant Potential. 96

Boiler Performance Factors 98

Boiler Dependability. 98

Economics of Boiler Operation 101

Flue Gas Quality. 105

Retrofit Effects. 118

Data Validity. 119

Quality Control 121

Specific Data Validity. 128

Backup Methods. 134

4. Special Projects. 135

Sulfate Formation Tests. 135

Purge Dryer Test Observations. 139

Benefits of Increasing Absorption Efficiency 139

Summary 139

5. Evaluation Methods. 146

Evaluation Goals 146

The Test Systems 147

Sensors and Analog Signal Transmission. 150

Sampling Systems. 152

Data Acquisition Systems (DAS). 157

Supplemental Data 160

Controlled Condition Test Methods. 162

Test Series One 163

Test Series Two 163

TABLE OF CONTENTS (continued)

Page

Test Procedures 166

Quality Control. 167

Data Analysis Methods. 168

Data Processing 168

Data Base and Analysis. 172

Flue Gas Flow Rate 178

References. 183

Appendices. 185

A. Backup Data 185

B. Boiler Description. 230

C. FGD Plant Description 247

D. Carbon Balance Flue Gas Flow Calculation. 257

E. Method for Calculating Sulfur Balance 270

F. Purge Dryer Observations. 275

G. Benefits of Increasing the Absorption Efficiency. 304

H. Performance Data Collected by NIPSCO (November 1979 - July 1980) 350

FIGURES

Number		Page
2-1	Block Diagram of Test System.	19
2-2	Block Flow Diagram of Major Process Steps	26
3-1	Effect of Heat Rate and Excess Air on Flue Gas Rates. . . .	34
3-2	Thirty-Day Average SO_2 Removal.	36
3-3	Cumulative Frequency Distribution of 24-hour Average SO_2 Removal.	37
3-4	Cumulative Frequency Distribution of One-Hour Average SO_2 Removal.	38
3-5	Maximum Sustainable SO_2 Feed Rates.	44
3-6	Boiler Load and Load Loss During FGD Operation.	46
3-7	Flue Gas Flow Rates at Actual Load and Load Potential . . .	47
3-8	FGD Reliability Index	55
3-9	FGD Plant Operating History	58
3-10	FGD Full Operation with Bypass Damper Open.	59
3-11	FGD Plant Limits to Utilization	61
3-12	57 to 80 MW Load Following Test	74
3-13	57 to 80 MW Load Following Test	75
3-14	80 to 57 MW Load Following Test	76
3-15	80 to 57 MW Load Following Test	77
3-16	Comparison Actual Steam Consumed as a Function of SO_2 Feed Rate with Design Value.	83
3-17	FGD Steam Consumption as a Function of Flue Gas Flow. . . .	85
3-18	Boiler Capacity Factor.	99

FIGURES (continued)

Number		Page
3-19	Comparison of Heat Rates with and without FGD Operation.	102
3-20	Gross Load as a Function of Boiler Heat Input During FGD Operation	103
3-21	Comparison of Net Load with and without FGD Operation . . .	104
3-22	Flue Gas Flow to Scrubber as a Function of Coal Rate and Scrubber Outlet O_2 Concentration	109
3-23	Effect of Air Inleakage on Flue Gas Flow Rate	112
3-24	Boiler Resistance as a Function of Coal Rate.	120
3-25	Correction Factor for DAS Gross Megawatts	125
4-1	Absorber Flow Diagram	138
5-1	Block Diagram of Test System.	148
5-2	Flue Gas Sampling System.	154
5-3	Flue Gas Sampling System.	155
5-4	Location of Inlet Flue Gas Sampling Position.	156
5-5	Evaluation Data Flow.	169
5-6	Boiler Load as a Function of Boiler Main Steam Rate	176
5-7	Correction Factor for DAS Gross Megawatts	177
5-8	CO_2 Concentration of Flue Gas as a Function of the Oxygen Concentration (Baseline II Coal).	182
B-1	Plan View of Boiler and FGD Plant	232
B-2	Mitchell No. 11 Boiler.	234
B-3	Fuel Distribution to Burners.	238
B-4	Boiler Load Ramp.	240
B-5	Mitchell Station - Unit No. 11.	241

FIGURES (continued)

Number		Page
B-6	Soot Blowers.	243
C-1	Schematic Diagram of FGD Plant.	250
D-1	Percent O_2 in Flue Gas by Volume, Dry Basis	268
D-2	Percent CO_2 in Flue Gas by Volume, Dry Basis.	269
F-1	Solution Feed Rate and Heat Duty During Purge Dryer Tests .	280
F-2	Motor AMP Range During Purge Tests.	281
F-3	Purge Dryer Capacity.	282
G-1	Wellman-Lord/Allied FGD Process	313
G-2	Effect of Liquor Flow Rate on Removal Efficiency.	321
G-3	Limits of Absorber Bypass to Meet NSPS of 1.2 lb $SO_2/10^6$ Btu	324
G-4	Effect on Stack Exhaust Temperature from Partial Treatment without Reheat for an Emission Rate of 1.2 lb $SO_2/10^6$ Btu	325
G-5	Effects of High Efficiency Absorption and Partial Flue Gas Treatment on Energy Consumption.	329
G-6	Effects of High Efficiency Absorption and Partial Flue Gas Treatment on Annualized Operating Costs.	334

TABLES

Number Page

2-1 Test Program Tasks. 5

2-2 Chronology of Events. 6

2-3 Plant Improvement Projects. 23

2-4 Demonstration Year 30-Day Operating Periods 29

3-1 Approximate Design Criteria of FGD Plant. 33

3-2 Comparison of SO_2 Removal for First and Second Demonstration Year. 39

3-2A SO_2 Emissions per Boiler Heat Input 40

3-3 FGD Plant Minimum Sustainable Load Test Results 42

3-4 Thirty Day Average and Sustainable SO_2 Feed Rates 43

3-5 Inlet Concentration of SO_2. 49

3-6 Definition of Viability Indices 52

3-7 Reasons for Interruption to Operations. 62

3-8 FGD Plant Startup Time. 63

3-9 Minimum Sustainable Load Test - Four Day Results. 66

3-10 Minimum Sustainable Load Test Comparison of Continuous and Manual Test Data. 67

3-11 Comparison of Operating Conditions at Absorber Turndown and Minimum Sustainable Load. 69

3-12 Maximum Sustainable Load Test 70

3-13 Purge Rates 79

3-14 Average Hourly Energy Consumption 82

TABLES (continued)

Number		Page
3-15	Effect of FGD Steam Consumption on Boiler	86
3-16	Effect of FGD Energy Consumption on Boiler	88
3-17	Energy Consumed as Natural Gas	88
3-18	Actual Annual Operating Cost - Second Demonstration Year	90
3-19	Capital Cost	91
3-20	Projected Annual Operating Cost	92
3-21	Escalation and Fluctuation of Raw Material and Byproduct Costs	92
3-22	Actual Annual Operating Cost - First Demonstration Year	93
3-23	Natural Gas Consumption	95
3-24	Byproduct Sulfur Production	96
3-25	Summary of Boiler Outages	100
3-26	Boiler Shutdowns	100
3-27	Additional Heat Input Requirements with FGD Operation	101
3-28	Coal Composition Summary	106
3-28A	Comparison of Coal Compositions for Controlled Condition Tests with Baseline II Tests	108
3-29	Variation of Excess Oxygen	111
3-30	Average Inlet Temperature	113
3-31	NO_x Emission Rate from Boiler	113
3-32	Particle Emission Rates and Concentration	115
3-33	Particle Emissions from Boiler: Baseline II and FGD Operation	116
3-34	Chloride in Coal During Periods of FGD Plant Full Operation	117

TABLES (continued)

Number Page

3-35 Steam Drain from Boiler by FGD Plant. 118

3-36 Instruments Used in Test Program, Serviced by Others. . . . 124

3-37 Uncertainty in Flue Gas Calculation 129

3-38 Sulfur Balance During Mode 2 Operation: May, June, July, and September. 130

3-39 Effect of Process Variables on Sulfur Inlet FGD Plant - Coal Method. 131

3-40 Month by Month Sulfur Balance 131

3-41 Comparison of Methods for Determining Sulfur Balance - July 132

3-42 Estimated Errors for Calculating Sulfur Balance Uncertainty. 133

4-1 Sulfate Formation Tests - Sampling and Analysis Schedule. . 136

4-2 Sulfate Formation Test Array. 137

4-3 Sulfate Formation Test Results. 140

5-1 Test Parameters 149

5-2 Test System Channels. 150

5-3 Continuous Analyzers for Flue Gas 152

5-4 Data Acquisition Components 157

5-5 Conversion Coefficient Values 158

5-6 Fly Ash Purge Analytical Methods. 161

5-7 Controlled Condition Test Program - Schedule. 164

5-8 Controlled Condition Test Program - Sampling and Analytical Schedule. 165

5-9 Flue Gas Sampling and Analytical Methods. 167

5-10 FGD/Boiler Operating Modes. 170

5-11 Assumed Values of Carbon Sources and Sinks. 179

TABLES (continued)

Number Page

A-1 - A-27 Backup Data 186

B-1 Fans and Pumps Information. 236

B-2 Regenerative Air Heater (Design Conditions) 245

F-1A Purge Dryer Heat Duty - Metric Units. 283

F-1B Purge Dryer Heat Duty - English Units 286

F-2 Purge Dryer Performance 289

F-3A Purge Dryer Test Data - Metric Units. 290

F-3B Purge Dryer Test Data - English Units 297

G-1 Operating and Cost Factors. 317

G-2 Maximum SO_2 Removal for Four-Stage Absorbers. 319

G-3 Cost Effects Within Maximum SO_2 Removal Range 320

G-4 Partial Treatment Limits for 115 MW Boiler. 326

G-5 Summary of Partial Treatment Effects on Exhaust Temperature and Stack Reheat Requirements. 328

G-6 Effect of Sulfur Content on Energy Consumption for Four State Operation and a Plant Employing Partial Treatment. 330

G-7 Boiler Derating Effects from High Efficiency Absorption and Partial Flue Gas Treatment 331

G-8 Comparative Cost Factors for Various Treatment Modes (115-MW Boiler, 3.15% Sulfur in Coal). 333

G-9 Annual Operating Costs for Four Stage Absorption and Partial Flue Gas Treatment 335

G-10 Relative Costs of Three-Stage and Four-Stage Operation. . . 337

G-11 Effect of Solution Rerouting. 339

TABLE OF CONTENTS (continued)

Number		Page
H-1	Hours Boiler Down or Unavailable: January-July	357
H-2	Viability Index Comparisons	358
H-3	Post Test Purge Rates	358
H-4	Raw Material and Byproduct Cost Comparisons	359
H-3.1	Data for 16 Days of Operation During the Period April 12-May 1, 1980	365
H-3.2(a)	FGD Viability Factors	366
H-3.2(b)	FGD Plant Operations Data	367
H-3.2(c)	Causes of FGD Down Time & Frequency	369
H-3.4(a)	FGD Plant Energy Consumption for 16 Days April 12 - May 1, 1980	370
H-3.4(b)	FGD Plant Energy Consumption January-July 1980	371
H-3.5(a)	FGD Operation Maintenance Costs - Dollars	372
H-3.5(b)	Economics - Soda Ash Consumption.	373
H-3.7(a)	Unit 11 Boiler Shutdown Data.	374
H-3.7(b)	FGD Plant Operating Hours	376
H-3.7(c)	Coal Analysis	383
H-3.7(d)	Flue Gas Temperatures	384

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EXECUTIVE SUMMARY

In 1972, the Environmental Protection Agency (EPA) entered into a cost-shared contract with Northern Indiana Public Service Company (NIPSCO) to design, construct, and operate a flue gas desulfurization (FGD) plant that uses the Wellman-Lord/Allied Chemical (WL/Allied) FGD process. NIPSCO entered into contracts with Davy Powergas (now Davy McKee) to design and construct the unit and with Allied Chemical to operate the plant. The FGD unit was retrofitted to NIPSCO's Unit No. 11 (Mitchell No. 11) boiler located at the D. H. Mitchell station in Gary, Indiana. The WL/Allied process developed by the two design organizations is a regenerable process that is based on the recovery of sulfur dioxide (SO_2) in concentrated form and its subsequent reduction to elemental sulfur. The product is sold to partially offset the process costs. This was the first coal-fired Wellman-Lord application, as well as the first joint Wellman-Lord/Allied Chemical installation.

To ensure that potential users are fully aware of the commercial practicality of the WL/Allied process, the EPA conducted a test program to evaluate the performance of the demonstration unit for a two-year period beginning in September 1977. This report analyzes the results of the test program during the two years of FGD plant operation. Operation of the FGD unit is continuing at this time.

The test program consisted of three major test phases:

1. A baseline test
2. Acceptance testing
3. Demonstration test and evaluation

The initial baseline and acceptance tests have been described in earlier reports.[1,2] The acceptance test for demonstrating guaranteed performance was successfully completed in September of 1977, and a scheduled one year of operation for demonstration testing followed immediately. The test program, as originally planned, was to include one year of testing and evaluation while the FGD unit was operating within the normal variation of boiler operation as well as out to the limits of boiler operation. However, the FGD plant achieved only 90 days of accumulated operation during the first year. It was apparent, after about six months of sporadic operation, that the FGD plant was not operating in a manner acceptable for commercial application due to factors not entirely attributable to FGD deficiencies. Two boiler related problems, coal feeding and boiler feedwater quality problems, contributed substantially to the poor utilization of the FGD plant and were not solved until the latter half of the year. Problems also were encountered at the boiler/FGD interface, in particular, booster blower and damper problems. It was decided that the major problems were probably correctable. Therefore, a plant improvement program was undertaken and targeted for completion during a scheduled boiler shutdown which coincided with the end of the one-year demonstration. Accordingly, it was decided to continue with the demonstration test program for another full year. The major improvements included boiler air preheater modifications and duct insulation to raise the flue gas temperature above the dew point, and reblading of the booster blower fan. The air preheaters were modified by removing part of the baskets used for heat transfer, thus raising the temperature of the flue gas entering the FGD plant. With this modification, and with more duct insulation, the inlet temperature averaged above 148.9°C (300°F) and there were no further problems at the booster blower due to wet flue gas. There was, however, some loss in boiler efficiency as a result of this change.

Test program results and operating experience during the second year showed a substantial improvement in FGD performance, and since boiler utilization was high, the results more nearly duplicated those expected during commercial application. The test results and operating experience for the second year of demonstration are summarized as follows.

Dependability

Reliability of the FGD unit (hours operated/hours called upon to operate) was 61%. However, FGD plant utilization was consistent for the last seven months and the reliability averaged 74% for that period. The reliability record was established with virtually no redundancy built into the FGD unit. Also, the evaporator was designed for the equivalent of only 80% of full boiler load. With limited surge capacity within the regeneration loop, the FGD plant was unable to effect complete SO_2 recovery during evaporator or reduction unit shutdowns. The FGD plant was utilized 56% of the time with the remaining time due to:

- FGD repair - 33%
- Startup and shutdown - 5%
- Boiler down - 6%

The SO_2 absorber was essentially trouble free. The reduction unit required the most downtime for repairs, followed by the evaporator circulating pump and the booster blower.

Sulfur Dioxide Removal

The process was controlled to remove 89% of the inlet SO_2. Sulfur dioxide removal efficiencies of 89% (24-hour average) or higher were accomplished 84% of the time. Removal efficiency was 85% or higher for 97% of the time. On a boiler heat input basis, SO_2 emissions were controlled in the range of 110 ng/J to 400 ng/J (0.25 $lb/10^6$ Btu to 0.94 $lb/10^6$ Btu).

Sulfur dioxide removal was attained at boiler loads in the range of 53 MWe to 85 MWe of the 115 MW boiler. The lower limit was set by the limiting turndown capability of the reduction unit. The upper limit was set by the capacity limitation of the evaporator, as designed. This would not have been a limitation had the evaporator been designed to match the full SO_2 removal capability of the absorber. Because a substantial amount of energy (primarily as boiler main steam) was consumed by the FGD plant, the generating potential of the boiler was actually about 95 MW at the FGD maximum capacity limit of 85 MWE. The term "generating potential" refers to the gross megawatts that the boiler is capable of generating during FGD operation but cannot attain due to the boiler main steam consumed by the FGD plant.

Design Limitations

The major design constraint was the design capacity of the recovery area which limited the boiler to 74% (85 MWe) of full load, if there was to be no loss of SO_2 removal efficiency (a summary description of the process is found in Section 2. More detailed descriptions of the boiler and of the FGD plant and process may be found in Appendices B and C). The test results confirmed this value as the upper limit of operation. Sulfur dioxide feed rates up to 0.68 kg/s (5,400 lb/h) were successfully treated by the absorber for sustained periods of at least 24 hours. Thus, design expectation for removal efficiency at a maximum SO_2 feed rate was met.

The SO_2 was fed at flue gas flow rates about 12% higher (at equivalent boiler load) than had been expected. However, the booster blower and the absorber were designed for a flue gas volume well in excess of the flue gas flow rates that were experienced. Performance at maximum design throughput could not be adequately tested because of the design limitations of the regenerating system. This was not a process limitation, and there was no reason to assume that additional capacity could not be designed into another unit to accommodate the total capacity of the boiler.

Turndown of the overall FGD unit was determined by the lowest practical operating limit of the reduction unit. Testing confirmed this to be at a load factor of 46% (53 MWe) and a SO_2 feed rate that was 63% of the maximum feed rate.

The purge treatment dryer achieved an actual capacity of about 50% of its design feed rate. This limitation was unexpected. It also should be correctable when designing another unit by redesign of the drying system, with a different dryer, or by using an antioxidant to limit the oxidation of the sodium sulfite contained in the absorbing solution to sodium sulfate.

Process Economics

The projected annual operating cost in 1978 dollars was 13.2 mills/kWh. Actual annualized cost for the second year of the demonstration, using 1978 prices for raw materials and utilities, was 14.9 mills/kWh. This is 13% higher than the projected unit operating cost. Byproduct sulfur production averaged 1.81 kg/s (17.3 ton/day). Credits for the sale of the sulfur amounted to only 0.2 mill/kWh. Actual annualized cost was based on the 82.2 MWe of generator output that was possible at the FGD design capacity of 92 MW, assuming 351 days of boiler operation per year. On this basis, the actual boiler capacity in total kilowatt hours was 92% of the projected capacity. Annualized costs are quite sensitive to lower than projected capacities because fixed costs, about 50% of the annual costs, and labor costs continue to accrue whether or not the boiler or the FGD plant is operating.

Energy and Raw Material Consumption

A significant amount of steam produced by the boiler was consumed by the FGD plant, required primarily by the evaporator to recover SO_2 and regenerate the scrubber solution. The same steam at higher pressures was used in steam turbine drives. Electrical energy consumption amounted to about one megawatt hour/hour after the evaporator circulating pump had been converted from steam turbine drive to electrical drive for improved operability. Actual steam consumption varied from 6.6 kg/s (52,000 lb/h) to 7.4 kg/s (59,000 lb/h) and averaged 5% higher than design expectations. In energy, this was equivalent to 11% of the boiler input energy derived from fuel. The equivalent loss in electrical generating capacity amounted to 8% (derating) of the boiler nameplate capacity of 115 MW at the average boiler load of 77 MWe. Including one megawatt of electricity consumed, the total energy requirement was 12% of the boiler heat input derived from fuel. This derated the boiler an additional 0.6%. The total derating is equivalent to an electric production loss of 10 MW of generator output.

Soda ash is used as makeup sodium carbonate (Na_2CO_3) for the scrubbing process. Makeup is made necessary by the buildup of inactive constituents in the absorber/evaporator loop, such as sulfate and thiosulfate, that must be purged. Any loss from the system due to leaks also would require soda ash makeup. High soda ash consumption during the first demonstration year was due to leaks at the bottom collector tray of the absorber that were repaired before commencing the second demonstration year. Average daily consumption of soda ash was 0.099 kg/s (9.4 tons/day), using the total operating days of the absorber/evaporator as the time base. The performance guarantee for acceptance was 0.069 kg/s (6.6 tons/day) at the design levels for flue gas flow and inlet SO_2.

Natural gas is used as the reductant for converting the SO_2 to elemental sulfur. It also is the fuel used to incinerate the tail gas emitted from the reduction process. The tail gas is returned to the inlet of the absorber after incineration. It was necessary to continue incinerator operation during shutdowns to destroy the reduced sulfur forms that desorb from the reduction unit refractory materials. Thus, there is a corresponding improvement in unit consumption of natural gas with improvement in reliability. About 0.2 m^3 (7 ft^3) of natural gas was consumed per pound of sulfur produced which is in accordance with the design expectations. During operation, the incinerator consumed 7.5% of the natural gas. In contrast, the incinerator consumed over 12% of the gas overall because it continued to operate during shutdowns, demonstrating the consequence of the 61% FGD plant reliability factor.

Purge Treatment Considerations

The purge unit, as initially designed, was to have treated a small purge stream removed from the regeneration loop, to effect a separation of sodium sulfate from most of the sulfite/bisulfite components, and to dry the sodium sulfate to produce a marketable product. The "wet" end of this purge treatment system performed satisfactorily. The limitation of the purge dryer already has been mentioned.

The amount of purge to be treated is a function of the formation of sulfate and possibly thiosulfate during absorption. Attempts to determine the amount of sulfate formation during absorption were frustrated by an inability to obtain correct flow measurements and uncertainties about the specific water balance across the absorber. However, the data seem to indicate that sulfate formation is a function of the oxygen concentration in the flue gas. Thus, higher than design purge rates might have been due to the high excess air levels in the flue gas. An average purge rate* for the last seven months of operation has been estimated at between 18.2% and 24.8%, substantially higher than expected. The estimate was determined from soda ash consumption and the calculated amount of SO_2 removed. A purge rate of about 10% was the value indicated during the design phase of the project. In conclusion, the information seems to indicate actual purge rates much higher than design, the magnitude of which is unclear. Davy McKee is investigating the use of an antioxidant that shows promise for reducing the oxidation rate and, thus, the amount of sodium values that must be purged as sodium sulfate.

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*Purge rate is the ratio of moles sodium in purge to moles SO_2 removed from flue gas, expressed as a percentage.

SECTION 1

INTRODUCTION

BACKGROUND

The Environmental Protection Agency (EPA) has undertaken a number of programs to demonstrate sulfur-oxide emission control processes that are applicable to stationary sources. These demonstration programs comprise operation of an emission control unit of such size and for such duration as to permit valid technical and economic scaling of operating factors to define the commercial practicality of the process for potential industrial users. One of the candidate processes being evaluated is the Wellman-Lord/ Allied Chemical (WL/Allied) process developed by Davy Powergas and Allied Chemical.

The Wellman-Lord sulfur dioxide (SO_2) removal process removes SO_2 from flue gas and recovers the sulfur values as SO_2, which then can be used to produce (by other processes) sulfur, sulfuric acid (H_2SO_4), or liquid SO_2. The Allied Chemical sulfur reduction process reduces the SO_2 to produce molten sulfur. The two processes were combined to demonstrate flue gas desulfurization (FGD) technology in which the scrubbing medium is regenerated and reused, and in which the SO_2 is recovered as sulfur, a marketable product. This configuration will be referred to as the WL/Allied process, although the two processes are not contingent upon each other and each can be used in other regenerable FGD configurations. A more detailed description of the FGD plant and process may be found in Appendix C.

In 1972, the EPA entered into a cost-shared contract with Northern Indiana Public Service Company (NIPSCO) to design, construct, and operate a FGD plant that uses the WL/Allied process. NIPSCO entered into contracts with Davy Powergas (now Davy McKee) to design and construct the unit, and with Allied Chemical to operate the plant. The FGD unit was retrofitted to NIPSCO's Unit No. 11 (Mitchell No. 11) boiler, located at the D. H. Mitchell station in Gary, Indiana.

The WL/Allied process developed by the two design organizations is based on the recovery of SO_2 in a concentrated form and its subsequent reduction to elemental sulfur. The product can be sold to partially offset the process costs. This was the first coal-fired Wellman-Lord application, as well as the first joint Wellman-Lord/Allied Chemical installation.

To ensure that potential users are fully aware of the commercial practicality of the WL/Allied process, the EPA conducted a test program to evaluate the performance of the demonstration unit. TRW was the test contractor. This report analyzes and discusses the test program results. It is the final report of a test program that began in 1973 with the development of test criteria and test plans and that was concluded in October 1979 when the last test data were collected.

PROGRAM GOALS

The principal objectives of the test program, as originally conceived, were as follows:

1. Verification of the reduction in pollutants achieved by the WL/Allied process FGD unit.
2. Validation of the estimated technical and economic performance of the demonstration unit.
3. Assessment of the applicability of the WL/Allied process to the general population of utility boilers.

The test program was designed to attain these objectives to the maximum extent possible. Emphasis on the various objectives was redirected due to test program findings and operating difficulties but, in general, the program goals remained unchanged throughout the program.

SECTION 2

PROGRAM OVERVIEW

CHRONOLOGY OF EVENTS

The major tasks of the test program are shown in Table 2-1. The occurrence of these events and how they have been coordinated with the tasks of the other project participants is set forth in this section. The order and timing of the significant milestones for the test program are shown in Table 2-2.

The test program was started in February 1973. At that time, the initial design phase for the FGD plant had been completed. The first task was to prepare a Work Plan Manual[3] that provided an overall plan for the entire test program. The initial test phase (July 1973) was a short test which characterized some of the properties of the flue gas emitted by the host boiler. Based on the information from these tests and from boiler and FGD design information, test criteria were established for the test program. These test criteria were included in a report that also included the characterization test results, a survey of potential users of the FGD process, and a survey of the marketing potential for the FGD byproducts. Concurrently, Davy McKee was developing detailed design and equipment specifications for the FGD unit. At the time, the design and construction was scheduled for an April 1975 acceptance test. However,

TABLE 2-1. TEST PROGRAM TASKS*

1. Prepare a Work Plan Manual.
2. Conduct a test to partially characterize the flue gas from the host boiler.
3. Prepare test criteria.
4. Conduct a review of fuel availabilities and of the market potential for the byproducts, sulfur and sodium sulfate (Na_2SO_4).
5. Prepare a test plan for baseline testing of the Mitchell No. 11 boiler.
6. Prepare a test plan for performance (acceptance) testing.
7. Prepare a test plan for testing and evaluation during a 1-year demonstration.
8. Design and install a test system.
9. Conduct the baseline test.
10. Conduct the performance test and confirm that the FGD plant has met its performance guarantees.
11. Conduct a study to determine the effects of improving the absorption efficiency of the Wellman-Lord adsorber.
12. Conduct a 1-year test to demonstrate the performance of the FGD plant (later extended to 2 years).
13. Conduct a second baseline test following the first year of demonstration testing.
14. Prepare reports, reporting and analyzing the test results.

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*The deliverables represented in this list are included in the list of references as References 1 through 9.

TABLE 2-2. CHRONOLOGY OF EVENTS

	Event	Date
1.	Start date.	February 1973
2.	Work Plan Manual completed.	June 1973
3.	Flue gas characterization tests run.	July 1973
4.	User Survey/Test Criteria report completed.	September 1973
5.	Baseline Test Plan completed.	December 1973
6.	Baseline Test I, first phase completed.	May 1974
7.	Acceptance Test Plan completed.	July 1974
8.	Baseline Test I, second phase completed.	April-May 1975
9.	Established test facility at D.H. Mitchell Station.	July 1975
10.	Demonstration Test Plan completed.	July 1975
11.	Baseline Test I report completed.	January 1976
12.	High efficiency scrubber study completed.	August 1977
13.	Acceptance test successfully completed.	September 14 1977
14.	Demonstration testing begins.	September 16 1977
15.	Flow confirmation test conducted.	October 1977
16.	Acceptance test report completed.	February 1978
17.	Demonstration first year completed.	August 1978
18.	Baseline Test II conducted.	October 1978
19.	Demonstration second year begins.	October 1978
20.	Demonstration second year ends.	October 1979
21.	Baseline Test II report completed.	November 1979

unforseeable delays later in the program prevented the meeting of that schedule.

Following establishment of test criteria, detailed test plans for a baseline test, an acceptance test, and a demonstration test and evaluation were completed. All test planning was completed by July 1975. The original schedule called for completion of the test planning by November 1974 but this schedule was extended while additional funding was being arranged to cover construction cost escalations. In 1974, a baseline test to fully characterize boiler performance was initiated but not completed due to boiler ID fan problems. The baseline test was completed in 1975.

Renegotiation for construction funding and equipment delivery delays resulted in the rescheduling of the acceptance test from an initial date of April 1975 to December 1975, with additional adjustments in the schedule through October 1976. Davy McKee began startup of the unit in the summer of 1976. Startup delays and a boiler shutdown resulted in further postponement of the acceptance test. In January 1977, a fire caused extensive damage to the boiler and the boiler was down for repairs until May 1977. Both the boiler and the FGD plant experienced some startup problems that delayed the acceptance test until August 1977.

During the construction and startup period and the associated delays, TRW was involved with installation of a test system and with a study to determine the effects of increasing absorber efficiency. A test facility was established at the NIPSCO site in July 1975. There was sufficient time to complete installation of the test system before the acceptance test.

The acceptance test was started on August 29, 1977 and was successfully completed on September 14, 1977. Following the acceptance test, the boiler was put on maximum load to establish that the booster fan and absorber could take all the flue gas at full load. A one-year demonstration and demonstration test program commenced on September 16. The FGD plant was shut down for two days in October to confirm whether or not flue gas flows were identical at baseline conditions and at FGD performance test conditions.

During the one-year demonstration, boiler and FGD operation was erratic. A program of boiler and FGD process improvements was started but had not been completed by the end of the year. This prompted a decision to continue with the demonstration for an additional year; therefore, demonstration testing was continued through October 1979. Following a scheduled boiler shutdown in September 1978, a second demonstration year was started in October. This was preceded by a second baseline test (Baseline II). NIPSCO has continued to operate the FGD plant following completion of the test program in October 1979. Performance data collected by NIPSCO for the period November 1979 to July 1980 were reported in Appendix H.

SUMMARY OF RESULTS

Demonstration Testing

The test program was designed to demonstrate guaranteed performance of the Wellman-Lord/Allied process, and its ability to meet these performance criteria in a long term, dependable manner relative to the specific flue gas conditions at the host site. Since the FGD plant was designed and sized for a specific load factor and specific flue gas characteristics, the test also evaluated its operability over the normal range of load variation and flue gas composition experienced during the demonstration period. The results are summarized as follows:

a. Reliability of the FGD unit, hours operated/hours called upon to operate, was 61%. The reliability record was established with virtually no redundancy built into the FGD unit. Also, the evaporator was designed for only 80% of full boiler load. Because of limited surge capacity within the regeneration loop, the absorption section of the FGD plant was not able to operate to effect complete SO_2 recovery during evaporator or reduction unit shutdowns.

b. The major sources of interruptions were:
   - The reduction unit
   - The evaporator circulating pump
   - The booster blower
   - The evaporator
   - Startup time

c. Twenty-four hour average SO_2 removal efficiencies of 85 to 92% were typical.

d. Sulfur dioxide removal was attained at boiler loads in the range of 53 MWe to 85 MWe of the 115-MW boiler. Some operation was achieved up to 93 MWe. The lower limit was set by the limiting turndown capability of the reduction unit. The upper limit was set by the capacity limitation of the evaporator as designed. Because a substantial amount of energy (largely as boiler main steam) was consumed by the FGD plant, the generating potential (actual megawatts generated plus MW equivalent of steam consumed by FGD plant) of the boiler was actually about 95 MW at the FGD maximum capacity limit of 85 MWe.

e. Sulfur dioxide removal was obtained from flue gas with the following characteristics relative to design:

- Sulfur dioxide feed in excess of the expected plant capacity of about 0.68 kg/s (5400 lb/h) was successfully treated for sustained periods of 24 hours and greater without loss of SO_2 removal efficiency. Overall for the second demonstration year, SO_2 feed averaged 0.59 kg/s (4700 lb/h).
- Flue gas flow rates were usually higher than the expected flow rate of 151 m^3/s (320,000 acfm) by a substantial amount. All flue gas flow rates in this report are at the design basis of 148.9°C (300°F) and 1 atm.
- Inlet temperature, following modifications to the air preheater of the boiler to obtain higher temperatures, averaged 151.6°C (305°F). Design basis temperature was 148.9°C (300°F).

f. The steam consumed by the FGD plant amounted to about 11% of the boiler input energy. Boiler derating averaged 9%.

g. Raw material consumption was as follows:

- Soda ash average daily consumption was 9,000 kg (9.9 tons). Moles of sodium consumed averaged 21.7% of the moles of SO_2 removed from the flue gas.

- Natural gas was consumed at a rate of 0.44 m^3/kg ($6.9 ft^3$/lb) of sulfur produced.

h. The production of sulfur as a byproduct averaged 0.182 kg/s (17.3 tons/day). The product was sold locally.

These results represent the performance during the second demonstration year, which was more typical of the expected performance. The detailed discussion and analysis of these results is presented in Section 3.

Baseline Testing

The first baseline test (Baseline I) was a comprehensive characterization of flue gas properties and their correlation with boiler operating factors. The data were used to show how some of the flue gas properties, particularly flow and temperature, differed from the parameter values used to design the FGD plant. The differences had a profound effect on the criteria for acceptance testing, and the baseline data were very useful for defining those criteria. The data also were useful for defining the range of testing.

Following completion of the first year of FGD process demonstration, baseline tests (Baseline II) were repeated with the following objectives in mind:

- Establish an updated boiler performance baseline for comparison with the boiler performance when the FGD plant is operating.
- Obtain an updated characterization of the flue gas leaving the boiler.

The collection of baseline data was necessary to establish to what extent a substantial rebuild of the boiler and a change in the type of coal being burned had affected the performance of the boiler. Much of the data was collected only to document non-retrofit boiler performance and provide a baseline for comparison with performance during FGD operation. Flue gas volumes, flue gas temperatures, and coal quality were of particular interest because of their impact on FGD operation and performance. The results are summarized below.

Flue Gas Volume--

Flue gas volumes were 27% higher than the volume of flue gas used for FGD plant design, 151 m^3/s at 148.9°C (320,000 cfm at 300°F). Measured values were confirmed by comparing them with calculated values that were determined by carbon balance. An excess of inleakage air appears to be the major contributor to the high flue gas volumes. Heat rates that were higher than the new boiler design heat rates resulted in excess of coal consumption and this probably had some effect. However, heat rates had not deteriorated since the first baseline tests. Factors contributing to the high heat rates were high turbine steam rates and, at 40% load, combustion air in excess of the operating set point. Turbine steam rates, in pounds main steam/kWh, averaged 7% higher than boiler design and were consistently high at all levels of load. Both the high steam rates and the high excess air add to the volume of flue gas that the FGD plant must treat.

The measured flue gas volumes were compared with estimated values obtained by a calculation procedure utilizing a carbon balance. Coal feed rate and composition and the carbon dioxide (or oxygen) concentration in the flue gas must be known. This estimating technique can be used to continuously monitor and record flue gas volume. During Baseline II, the flue gas volumes using the estimating technique were in acceptable agreement with the measured values.

Flue Gas Temperature--

Flue gas temperature averaged 148.9°C (300°F) at all levels of load during Baseline II testing. This was substantially higher than the temperatures measured during Baseline I. The higher temperatures were due to the removal of part of the heat transfer surfaces of the air preheaters in a deliberate attempt to raise the temperatures above the dew point and thus prevent the scaling and corrosion that was occurring at the FGD booster fan. The temperatures were well above the sulfuric acid dew point.

Coal Quality--

Captain coal, the high sulfur (~ 3% sulfur) coal selected for use during FGD operation, was burned during Baseline II and during subsequent demonstration testing. The coal sulfur content during Baseline II was 0.2 to 0.3 wt.% lower than the Baseline I sulfur content. Except for the expected differences in variability of the ash and moisture contents, the quality of the two coals were about the same.

Sulfur Oxides (SO_x) Emissions--

Sulfur dioxide in the flue gas averaged 6% higher than the FGD design values, on a mass rate basis; however, closure of the sulfur balance across the boiler was more than 100%. Thus, depending on the source of the apparent error in the sulfur balance, the actual SO_2 feed rates may be close to or no different from FGD design. The SO_2 in the flue gas averaged 0.6488 kg/s (5,149 lb/h).

Acceptance Testing

Process performance guarantees were met or exceeded as confirmed by the acceptance testing which began on August 29 and ended on

September 14, 1977. The acceptance test consisted of two test periods. The design load test period was to be a 12-day period during which the FGD plant was to be operated at the design condition of a boiler flue gas output rate equivalent to 80% of the maximum boiler load of 115 MW gross. The FGD plant was operated for 265 hours at this load condition. The high load test period was to be an 83-hour period during which the FGD plant treated flue gas volumes equivalent to 95% of maximum boiler load. Actual operating time was 86 hours at this load condition.

Specific performance criteria were met or exceeded as follows:

a. Sulfur dioxide removal of 90% or better was achieved for 261 hours of the 265 hours of operation at design load conditions and was achieved for 84 hours of the 86 hours of operation at high load conditions.

b. Particulate emissions did not exceed 40 ng/J (0.1 lb/10^6 Btu) of boiler heat input at either design load or high load conditions.

c. The consumption of steam, natural gas, and electrical power averaged 76% of the performance guarantee requirements at design load conditions.

d. Soda ash consumption averaged less than 0.069 kg/s (6.6 tons of sodium carbonate [Na_2CO_3] per day) which was the limit set in the performance guarantees that was not to be exceeded during the design load period.

e. Sulfur product purity was greater than 99.5% at both design load and high load conditions. As a check to ensure that the sulfur product was acceptable for burning in a contact acid plant, impurities in the sulfur were compared with bright sulfur purchase specifications. All impurities were less than specified.

During the Design Load period, a SO_2 removal efficiency of 90% or better, based on 2-hour averages, was achieved for all but 4 hours of the 265-hour operating period. Sulfur dioxide removal was 88% and 89% (two-hour averages) for two 2-hour periods. Four hours were added to the 12-day test period to bring the total time of compliance operation to 12 days. During the high load period, there was one 2-hour period during which SO_2 removal was 89%. Similarly, three hours were added to the 83-hour test period.

TEST PLANNING

The test program as originally designed consisted of three major test phases:

1. A baseline test
2. Acceptance testing
3. A demonstration test and evaluation

Specific tests were designed for each of these test phases. However, the test designs were preceded by a preliminary work plan and test criteria development that defined the objectives and the scope of the test program and which also developed some preliminary test plans for budget purposes. The acceptance test design differed from the other designs because it focused entirely on conformance with the performance guarantees. The other essential criteria were careful quality control, backup in case of measurement equipment failure, and immediate reporting of the results.

The baseline test plan was interactive with the demonstration test plan in three respects:

1. The baseline test was to define the independent variables that potentially affect the performance of the FGD unit, and to determine the limits of these variables.
2. The baseline test was to define the performance range of the host unit as a benchmark that other potential users could use for comparison.
3. Baseline data was collected to determine the effects of retrofitting.

These were the major criteria used to design the baseline test. A demonstration test had to be designed to match these criteria in addition to the overall program objective of demonstrating the pollutant removal, economic performance, and technical performance of the WL/Allied process. With these criteria in mind, a comprehensive demonstration test program was designed.

The demonstration test and evaluation was a one-year continuous test program which included intensive shorter-term supplemental test periods (controlled condition tests). The test program utilized high frequency time-based data to develop performance information on the most typical operating conditions of the FGD plant. Flue gas parameters, boiler operating parameters, and selected performance parameters were monitored without interruption during the demonstration year. Other periods were allocated to conduct additional tests consisting of:

- Tests for collection of data not amenable to continuous monitoring.

- Tests at specified normal operating conditions of Mitchell No. 11 (controlled conditions).
- Tests at specified boiler operating conditions that were not normal for Mitchell No. 11.
- Tests utilizing manual sampling and analytical techniques for the measurement of flue gas parameters.

A monitoring system was provided for the continuous monitoring of as many as possible of the parameters of interest. The system employed continuous monitoring and a data acquisition system for scanning the sensor outputs at three- or six-minute intervals to develop a data base of one-hour averages. Where continuous monitoring was impossible, data was collected at lesser frequencies or intermittently. The one-hour averages supplemented with the data collected by low frequency sampling were processed by computer.

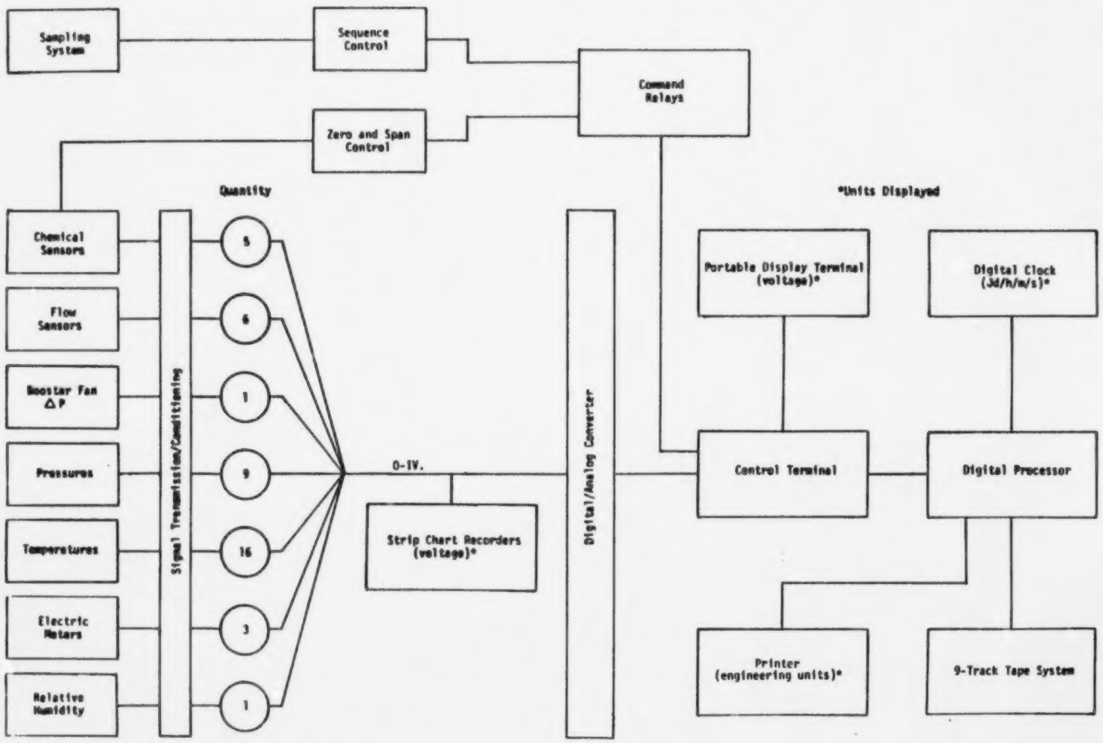
The data were evaluated in response to the objectives of the test program, with major emphasis on evaluating the pollutant reduction performance of the WL/Allied process. The data evaluations consisted of:

1. An assessment of the response of the major dependent variables to changes or fluctuations in the primary independent variables.
2. A demonstration of trends and cyclical patterns.
3. A documentation of operating costs and of the reliability of the FGD plant.
4. A documentation of the operating performance of Mitchell No. 11 before and after retrofit, using the results from baseline testing for comparison.

Results and interpretations were incorporated in a comprehensive report at the completion of the test program. The results also were summarized in reports submitted periodically during the demonstration year.

Specifying, procurement, and installation of the test equipment were essential elements of the test planning. A substantial amount of used, government-owned equipment was effectively utilized at some cost savings. Detailed specifications were written for equipment that had to be purchased. The two major equipment items purchased were the SO_2 analyzer and associated sampling system and the data acquisition system (DAS). Two trailers were leased for housing the test system and for a storage, laboratory, and coal handling area. Upon delivery, the equipment was assembled and checked out at test contractor facilities. After establishing a test site facility, the trailers and equipment were permanently located at the D. H. Mitchell station and sample lines and signal wire were run to the various sampling positions.

The test system (Figure 2-1) consisted of sensors for various boiler and flue gas operating variables (with emphasis on the FGD inlet and outlet flue gas parameters) and the accumulation of sensor analog signals by the DAS. The frequency of analog signal scan by the DAS was three or six minutes and data derived from these scans were used to compute one-hour averages. The DAS had the capability for storing the data on magnetic tape; however, hardware difficulties with the tape transport unit were experienced and thus very little automated data reduction was possible. Backup storage was available on teletype printouts or on charts taken from strip chart recorders.



Sampling System
Sequence Control
Zero and Span Control
Command Relays
Quantity
*Units Displayed
Chemical Sensors
Flow Sensors
Booster Fan ΔP
Pressures
Temperatures
Electric Meters
Relative Humidity
Signal Transmission/Conditioning
5
6
1
9
16
3
1
0-1V.
Strip Chart Recorders (voltage)*
Digital/Analog Converter
Portable Display Terminal (voltage)*
Digital Clock (Jd/h/m/s)*
Control Terminal
Digital Processor
Printer (engineering units)*
9-Track Tape System

Figure 2-1. Block diagram of test systems.

THE TWO-YEAR DEMONSTRATION

Sections 3 and 4 will focus on the results of the demonstration test and evaluation, particularly on the second year of the demonstration. Baseline and acceptance test results have been reported.[1,2,9]

The original test program was to include one year of test and evaluation during the year immediately following the acceptance test. It became apparent after about six months of sporadic operation that the FGD plant was unable to operate in a manner acceptable for commercial application due to factors not entirely attributable to FGD deficiencies. During a midyear review it was concluded by the project participants that the major problems were either boiler related or were problems encountered at the boiler/FGD interface, in particular, booster blower and damper problems. At that time it was decided that the major problems probably were correctable and to do this a plant improvement program was initiated. The improvements were to be substantially completed before the end of the scheduled boiler outage in September 1978. The boiler outage coincided with the end of the demonstration year. The test program was continued for another 12 months beginning with boiler startup following the scheduled shutdown in September. Test results for the first demonstration year have been reported.[8] These results are summarized below.

The FGD plant had a reliability factor of 50% (hours operated/hours called upon to operate) despite only 90 days of total operation. Reliability is the ability of the FGD plant to operate within specific limits of boiler operation. There was considerable operation of the boiler in an off-normal condition as a result of coal feeding and boiler feedwater problems. Overall, the boiler was operated for a total of 325 days of the year, but for only 179 days at stable enough conditions for operation of the FGD plant.

The SO_2 removal performance guarantee of 90% was met or exceeded 45% of the time, based on one-hour averaging times. The average removal efficiency was 89% and was met or exceeded 66% of the time. The operating set point was used to maintain a 90% reduction in the SO_2 concentration on a wet volume basis. This equates to about 89% removal after the dilution effects resulting from added water in the flue gas are considered.

Economic performance during the demonstration year was distorted by considerable off-normal operation of the boiler which limited utilization of the FGD plant, and by partial operation of the FGD plant (not counted as operating time) during which utility and raw material costs continued to accrue. The annualized unit cost of operating the FGD plant amounted to 15.8 mills/kWh compared with a projected annual unit cost of 14.9 mills/kWh. The high costs despite the low utilization of the FGD plant reflect fixed charges and standby operating costs such as labor.

The effect on boiler operation from the FGD installation was threefold. First, substantial electric power was not available for distribution as a result of FGD plant energy usage, primarily as steam. During a 42 day sustained run of the FGD plant, the unavailable power amounted to nearly 11 MW. Second, the boiler was limited to a sustained load of 92 MWe by FGD capacity limitations. Operation at higher loads was possible for only limited periods of time. During the same 42 day run, which followed correction of the coal feeding and the water quality problems, boiler gross output averaged 79 MW while the FGD plant was operating. Without the FGD plant, the boiler could have generated 89 MW of electric power with the same heat input. Third, there was also a lower limit of operation below which the SO_2 reduction unit would not operate. This established minimum limits for boiler load or for coal sulfur content. The lower limit was established during the second demonstration year.

About midway through the first demonstration year, booster blower problems prompted the initiation of a series of improvements to minimize FGD down time. The major improvements included boiler air preheater modifications and duct insulation to raise the flue gas temperature above the dew point, and included reblading of the booster blower fan. Because these projects could not be completed before the end of the demonstration year, the test program was continued for another year. The plant improvements program was initiated in June 1978 and included several projects for the purpose of eliminating or minimizing various operating problems. The projects and approximate completion times are presented in Table 2-3. These modifications were at least partially responsible for some of the improvements in boiler and FGD plant performance that are reported in Section 3.

The follow-on test program was essentially an extension of the first year of testing and evaluation; however, emphasis was placed on the reliability of FGD plant operation while following normal swings in boiler load. In addition, controlled condition tests were conducted to evaluate the FGD system at its capacity limits and to establish the load

TABLE 2-3. PLANT IMPROVEMENT PROJECTS

Item	Date completed	Action
Coal supply	Completed June 78	An uninterrupted supply of Captain coal available for Mitchell No. 11 use.
Air heater	During September shutdown	Part of baskets which provide heat storage removed to raise inlet duct temperature.
Duct insulation	After September shutdown	Insulate duct before and after booster blower.
Blanks	During September shutdown	Provision to install blanks rapidly at inlet of booster fan as an alternative to the isolation damper.
Booster blower	Not completed	Install a sparger pipe in the booster blower to periodically steam clean blades while in run.
Evaporator pump	During September shutdown	Install an electric motor as an alternative to steam turbine drive.
Absorber	During September shutdown	Recoat and repair leaks.
Booster blower turbine	After September shutdown	Provide enclosure to protect against SO_2 and weak acid attack.
Sulfur condenser	During September shutdown	Plug leaking tubes.

following capability of the FGD unit. Other nonroutine testing was done to determine the sulfate formation rate during SO_2 absorption. Data also was compiled on tests that Davy McKee conducted to determine the capacity of a purge treatment dryer. As part of the follow-on program, boiler baseline data were collected with the FGD plant down and completely isolated (Baseline II). The results were to show any differences in boiler operating characteristics compared to the results of Baseline I performed before the installation of the FGD plant.

To help clarify the results of the second demonstration year, brief descriptions of the FGD process and of the performance evaluation methods employed are presented below. More detailed descriptions of the boiler and of the FGD plant and process may be found in Appendices B and C.

Process Description

Flue gas from Mitchell No. 11 is delivered to the suction of the FGD plant's booster fan. Mitchell No. 11 is a 115-MW pulverized coal-fired, balanced draft boiler with cold end electrostatic precipitator (ESP) particle control. The boiler was designed to use a coal with a nominal sulfur content slightly above 3%. The FGD unit was designed to accept flue gas at SO_2 concentrations equivalent to a 3% sulfur level in the coal.

The WL/Allied FGD process removes SO_2 from the flue gas stream by scrubbing with an aqueous sodium sulfite/bisulfite solution and subsequent thermal regeneration to recover the SO_2. The liberated SO_2 is then reduced to elemental sulfur which is sold. The FGD unit was designed to remove 90% of the SO_2 delivered with the flue gas at flue gas rates equivalent to a boiler load of 92 MW (80% of full boiler load). The absorber was designed to take up to about 183 m^3/s (388,000 acfm) (110 MW equivalent) of flue gas but this rate can be sustained for only a limited period of time because of the limited capacity of the solution regeneration portion of the FGD plant.

The critical design criteria are approximately as follows:

Flue gas temperature	300°F	148.9°C
Flue gas pressure	14.7 psia	101 kPa
Maximum flue gas flow	388,000 acfm	183 m^3/s
Gross MW equivalent	110 MW	110 MW
Steam equivalent	749,000 lb/h	94.4 kg/s
Design flue gas flow	320,000 acfm	151 m^3/s
Gross MW equivalent	92 MW	92 MW
Steam equivalent	603,000 lb/h	76.0 kg/s
Inlet SO_2 at design flow	4,842 lb/h	0.6101 kg/s
Equivalent SO_2 concentration	2,185 ppmv	2,185 ppmv

Any combination of flue gas volume and inlet SO_2 concentration that results in an SO_2 feed rate greater than about 0.6 kg/s (5,000 lb/h) for periods up to 83 hours is excess capacity for the recovery area. This means that sustained operation at excess capacity would lower the performance level to below 90% SO_2 recovery. The absorber and the recovery area were designed to have the turndown capability for steady state operation down to 46 MW boiler load. However, the lower limit for sustained operation of the reduction area is higher than 46 MW due to the operating characteristics of the reduction system.

Figure 2-2 shows the process steps. The FGD plant accepts the total flue gas stream from the discharge of the boiler's induced draft

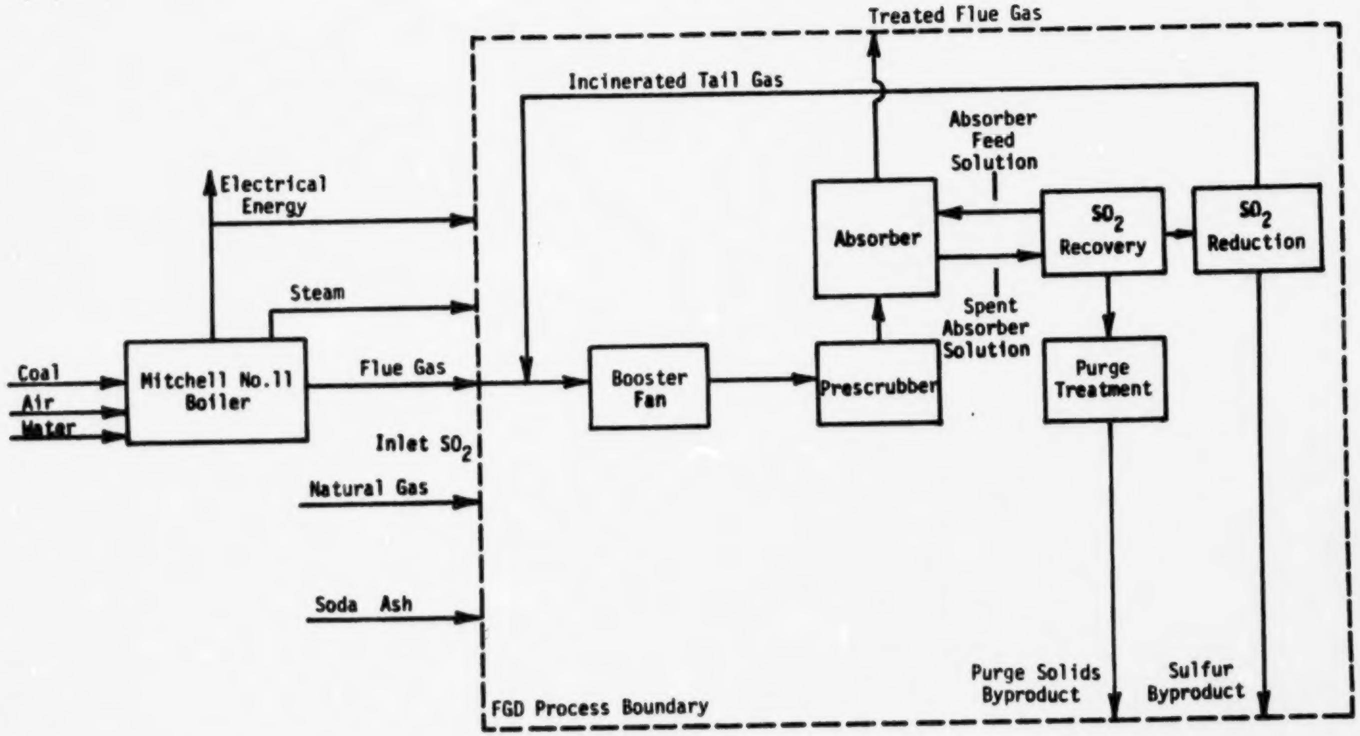

Treated Flue Gas
Incinerated Tail Gas
Absorber Feed Solution
Electrical Energy
Absorber
SO_2 Recovery
SO_2 Reduction
Steam
Spent Absorber Solution
Coal
Air
Water
Mitchell No.11 Boiler
Flue Gas
Booster Fan
Prescrubber
Purge Treatment
Inlet SO_2
Natural Gas
Soda Ash
Purge Solids Byproduct
Sulfur Byproduct
FGD Process Boundary

Figure 2-2. Block flow diagram of major process steps.

(ID) fans using a booster fan to force the flue gas stream through the prescrubber and absorber. The prescrubber is a single-stage orifice contactor that removes particulate matter in addition to that removed by the boiler's ESP. A pump recirculates the scrubber water from a sump back to the contactor. In order to control solids buildup in the liquid stream, a purge stream is withdrawn and makeup water is added to the prescrubber to compensate for this loss as well as to humidify the flue gas. The purge stream is sent to the power station's fly ash settling ponds.

The cooled, humidified flue gas leaves the prescrubber and enters the bottom of a three-stage absorber where the gas is contacted with the sulfite/bisulfite solution flowing countercurrently to the gas stream. The solution absorbs the SO_2 and the treated flue gas is then discharged to the atmosphere through a stack.

The spent sulfite/bisulfite solution is removed from the bottom tray of the absorber and sent to a surge tank for storage prior to regeneration in the SO_2 recovery step. To recover the SO_2, the spent absorbent is regenerated in a steam-heated, single-effect evaporator and then is returned to the absorber feed tank. The surge tank and absorber feed tank provide surge capacity for operating, for limited time periods, at flue gas rates in excess of 92 MW equivalent. To prevent the accumulation of sodium sulfate in the absorbing solution stream, a purge stream is sent to the purge treatment area. Here, the purge stream is crystallized and centrifuged and the solid product is removed and dried, yielding a salable sulfate byproduct. The sodium values lost in the purge stream are made up by adding sodium carbonate (Na_2CO_3) to the regenerated sulfite/bisulfite solution.

Sulfur dioxide released in the evaporator is transported to the SO_2 reduction area. The reduction step is a proprietary process developed by Allied Chemical which utilizes natural gas (CH_4) for the reduction of SO_2 to hydrogen sulfide (H_2S) and, ultimately, to elemental sulfur in molten form. A small stream of tail gas is returned after incineration to the inlet of the booster blower.

Performance Evaluation Methodology

Evaluation was in response to the test objectives and proceeded in six steps:

1. Collect applicable data and operating information.
2. Define hours of operation within each operating mode.
3. Process the raw data for each consecutive 30-day period and for specific periods according to the mode of operation.
4. Assess performance with regard to pollutant removal, dependability, energy consumption, and costs.
5. Assess the response of selected dependent variables to changes or fluctuations in the major independent variables.
6. Assess the effect of upsets and transients on SO_2 removal capability.

The methods employed made maximum use of the continuous monitoring of primary evaluation parameters. The continuous data, consisting of 26 boiler and FGD plant operating and performance parameters, were sampled every six minutes by the DAS. One-hour averages were computed by the DAS. Software developed specifically for this test program was employed to analyze the data by way of a two-phase computer program. The first phase consisted of a listing of hourly averages of the measured and calculated parameters needed as a data base. The second phase was generated from the data base and consisted of data summaries that identified the limits, central tendencies and variabilities in the data.

Continuous monitoring did not provide all of the needed data. For example, coal composition had to be determined on periodic spot samples and natural gas flow was read from flow totalizers only once each day. Sulfur production and dry purge rates were obtained from Allied once each month. Data of this type was incorporated into the data base.

The data was processed by computer to obtain a data base and a data summary for each 30-day period. All months were 30 days long and the periods were assigned to conform as closely as possible to calendar months. There were thirteen 30-day periods in the second demonstration year, including time for the Baseline II testing (Table 2-4).

TABLE 2-4. DEMONSTRATION YEAR 30-DAY OPERATING PERIODS

Calendar month	Start to end (0800 to 0800)
October 1978	10/04/78 - 11/03/78
November	11/03/78 - 12/03/78
December	12/03/78 - 01/02/79
January 1979	01/02/79 - 02/01/79
February	02/01/79 - 03/03/79

(continued)

TABLE 2-4 (Continued)

Calendar month	Start to end (0800 to 0800)
March	03/03/79 - 04/02/79
April	04/02/79 - 05/02/79
May	05/02/79 - 06/01/79
June	06/01/79 - 07/01/79
July	07/01/79 - 07/31/79
August	07/31/79 - 08/30/79
September	08/30/79 - 09/29/79
October	09/29/79 - 10/29/79

Two terms are used repeatedly throughout Section 3 and require definition. One of the terms is operating "mode". There were various modes of operation depending on the availability of equipment and on flue gas availability. The principal operating modes were as follows:

Mode 1 - Boiler operating, FGD not operating.

Mode 2 - Integrated operation of the absorber/evaporator and reduction units without bypass of any of the flue gas.

Mode 3 - Operation of absorber/evaporator with the flue gas bypass damper open. Bypassing of the flue gas may or may not occur depending on the booster blower speed setting.

Mode 4 - Operation of absorber/evaporator with the reduction unit not operating. Recovered SO_2 (from evaporator overhead) is vented after dilution with flue gas from other units.

Analysis of the data focuses on Mode 2 operation. The other term that needs to be defined is "generating potential." This term refers to the gross megawatts that the boiler is capable of generating during FGD operation but cannot attain due to the boiler main steam consumed by the FGD plant. Generating potential is determined by computing the megawatts equivalent to the FGD steam consumption and adding this value to the megawatts actually generated.

SECTION 3

DISCUSSION OF RESULTS

The second demonstration year was to cover 13 months, beginning on October 4, 1978 and ending on October 29, 1979. Startup of the FGD plant was delayed to conduct the Baseline II testing, thus, full operation of the FGD plant was not achieved until October 19, 1978. Testing ended on October 23, 1979 when the boiler went down due to turbine blade failures.

The FGD plant operated with only minor interruptions until December 5, 1978. From December 1978 through April 1979, operation was sporadic and from April through October 1979 a more consistent pattern of operation was established. Because of this consistency and the resulting improvement in plant reliability, much of the data analysis has been based on the results obtained during the April to October period. Additionally, results during the month of July are often cited because they include a wide range of boiler loads and because of the relatively high reliability obtained during that month.

SULFUR DIOXIDE REMOVAL

The performance guarantees specified 90% removal of the SO_2 at the inlet design conditions shown in Table 3-1. During baseline and acceptance testing, it was found that flue gas flows at 92 MW were

TABLE 3-1. APPROXIMATE DESIGN CRITERIA OF FGD PLANT*

Parameter, units	Quantities
Flue gas temperature, °C (°F)	148.9 (300)
Flue gas pressure, kPa (psia)	101.41 (14.7)
Maximum flue gas flow, m^3/s (acfm)	183 (388,000)
Gross MW equivalent, MW	110 (110)
Steam equivalent, kg/s (lb/h)	94.4 (749,000)
Design flue gas flow, m^3/s (acfm)	151 (320,000)
Gross MW equivalent, MW	92 (80% load)
Steam equivalent, kg/s (lb/h)	76.0 (603,000)
Inlet SO_2 at design flow, kg/s (lb/h)	0.6101 (4,842)
Equivalent SO_2 concentration, ppmv	2,185
Sulfur in coal, wt.%	3.16

*Unless otherwise indicated, all flue gas flows in this report are corrected to 148.9°C (300°F) and 101.3 kPa (29.92 in Hg).

significantly higher than 151 m^3/s (320,000 acfm). Figure 3-1, reproduced from the Baseline II report, shows that the calculated volume of flue gas at 92 MWe was almost 12% greater than design as a result of higher values of actual excess air than the design basis excess air of 5.6 vol.% oxygen (O_2), dry, at the inlet of the FGD plant. The actual excess oxygen was 7.4 vol. %. Baseline II tests were performed with Captain coal, the same coal used during the second year of demonstration testing. With a higher flue gas volume than design, a lower inlet SO_2 concentration may be expected at a coal sulfur content of 3.15 wt.%.

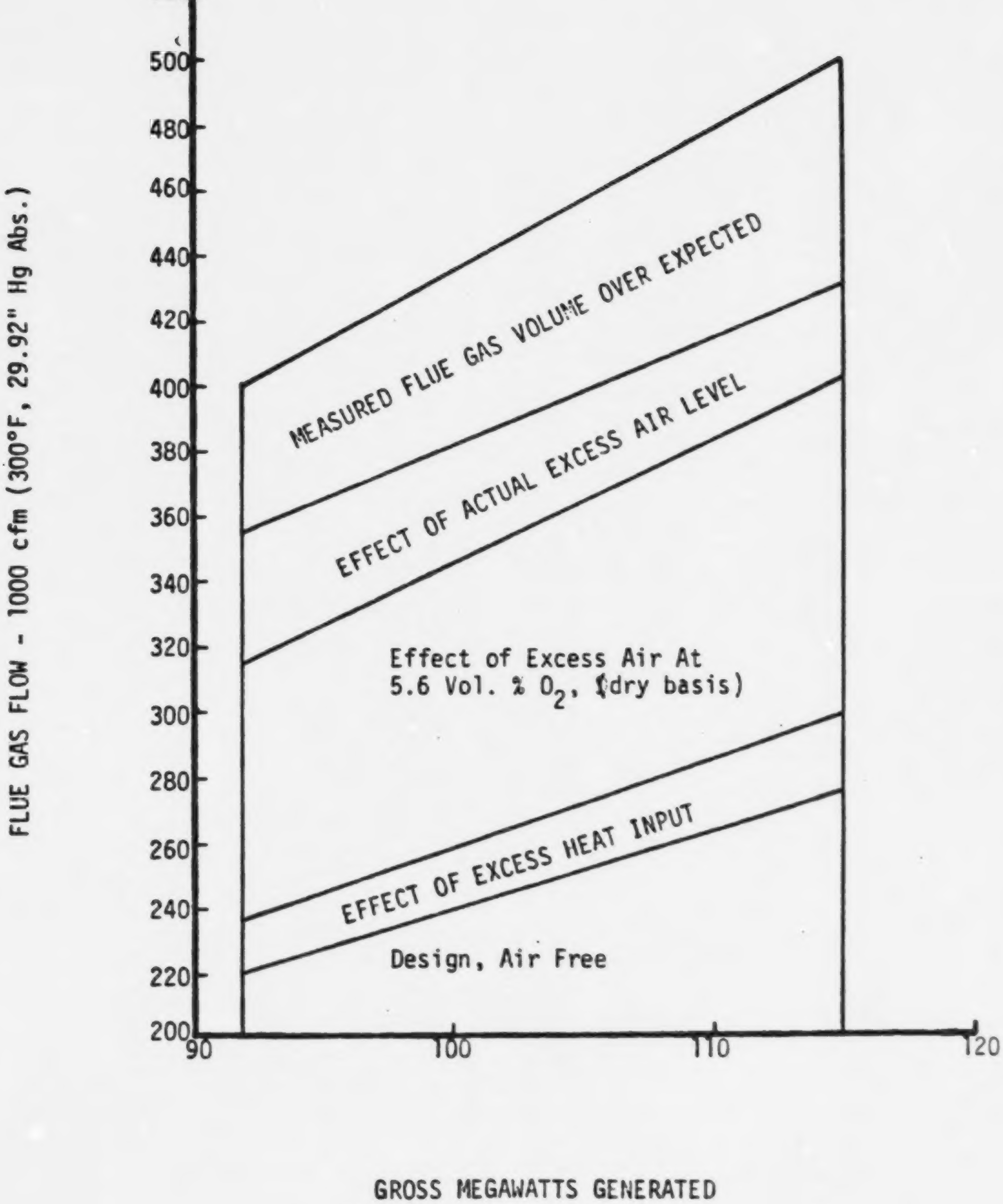


Figure 3-1. Effect of heat rate and excess air on flue gas rates.

However, we are more concerned with the measured mass rate of SO_2 and how it compares with the design value of 0.6101 kg/s (4,842 lb/h). For analyzing SO_2 removal performance, an upper design limit of 0.6779 kg/s (5,380 lb/h) is projected from the design basis coal sulfur limit of 3.5 wt.%. Also, the recovery unit was designed to process the sulfur removed at 80% of full boiler load, and this tended to further restrict the operating range during evaluation of SO_2 removal performance.

Removal Performance

Removal efficiencies averaged 89.6% overall for 11 of the 13 months of operation. January and February were excluded because of the limited amount of Mode 2 operation. Thirty-day average removal efficiencies varied from 88% to 93% (Figure 3-2). The range of values increases as averaging time decreases. Figures 3-3 and 3-4 are frequency distributions of 24-hour averages and 1-hour averages, respectively. These data show that SO_2 removal performance compared to design (90%) and to the operating control point of 89% was as follows:

	Percent of time operated	
Sulfur dioxide	24-h average	1-h average
90% and greater	60	52
89% and greater	84	78
85% and greater	97	97

Removal efficiencies of less than 85% occurred infrequently. There are records of three 24-hour periods that showed SO_2 removals of less than 85%:

1. April 2 (82%) - no explanation of problem.
2. August 23 (81%) - no explanation of problem. Inlet SO_2 was much lower than normal.

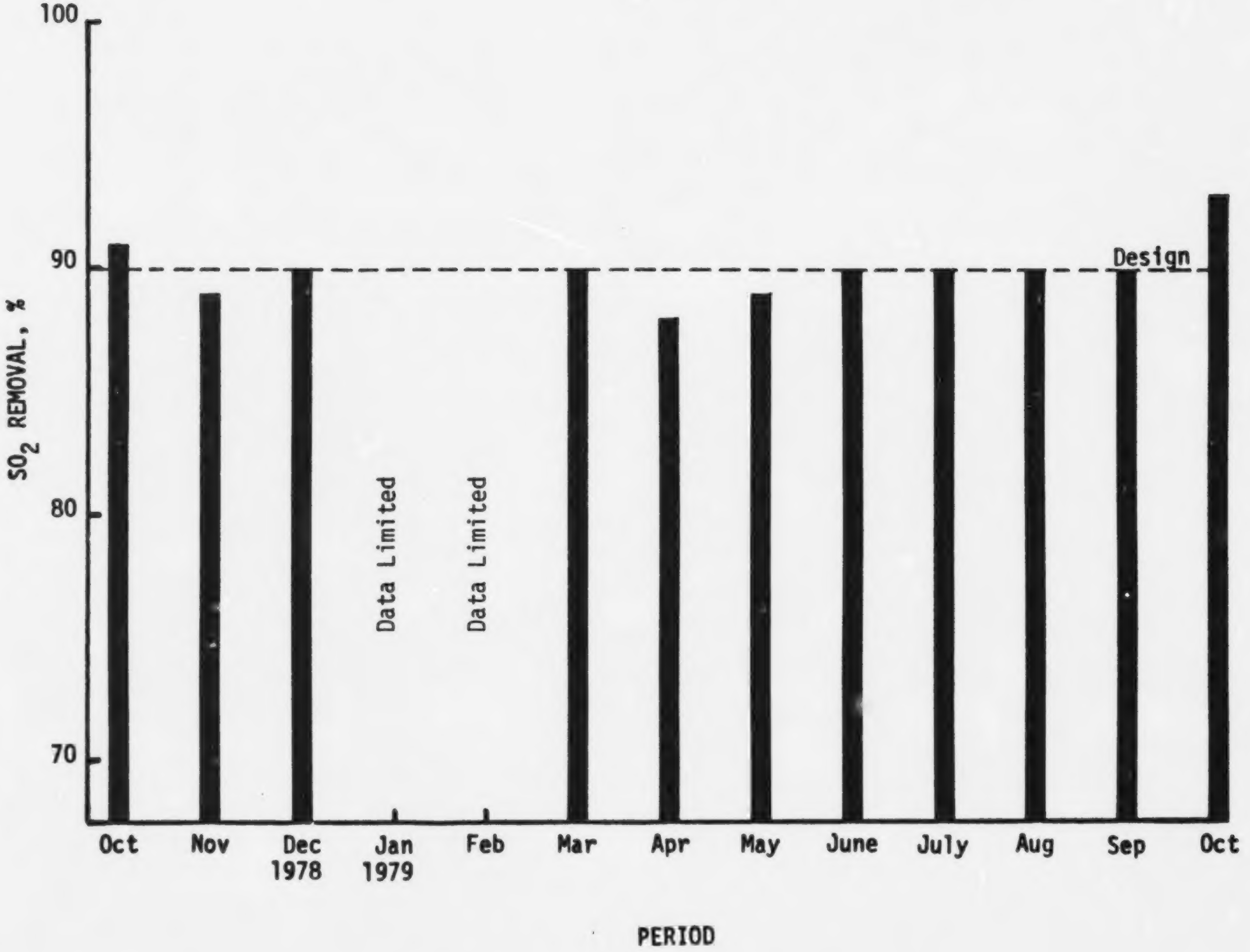


Figure 3-2. Thirty-day average SO_2 removal.

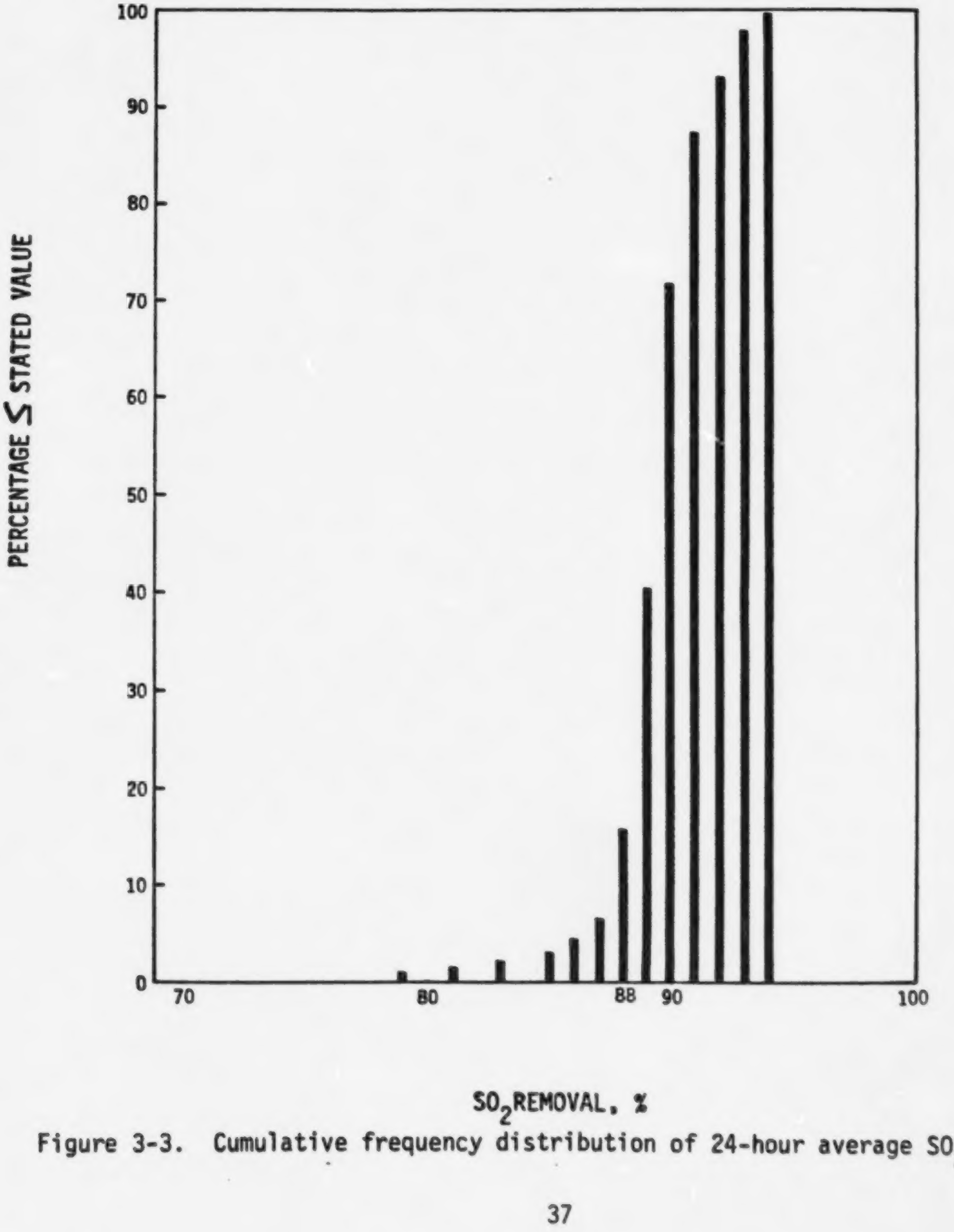


Figure 3-3. Cumulative frequency distribution of 24-hour average SO_2 removal.

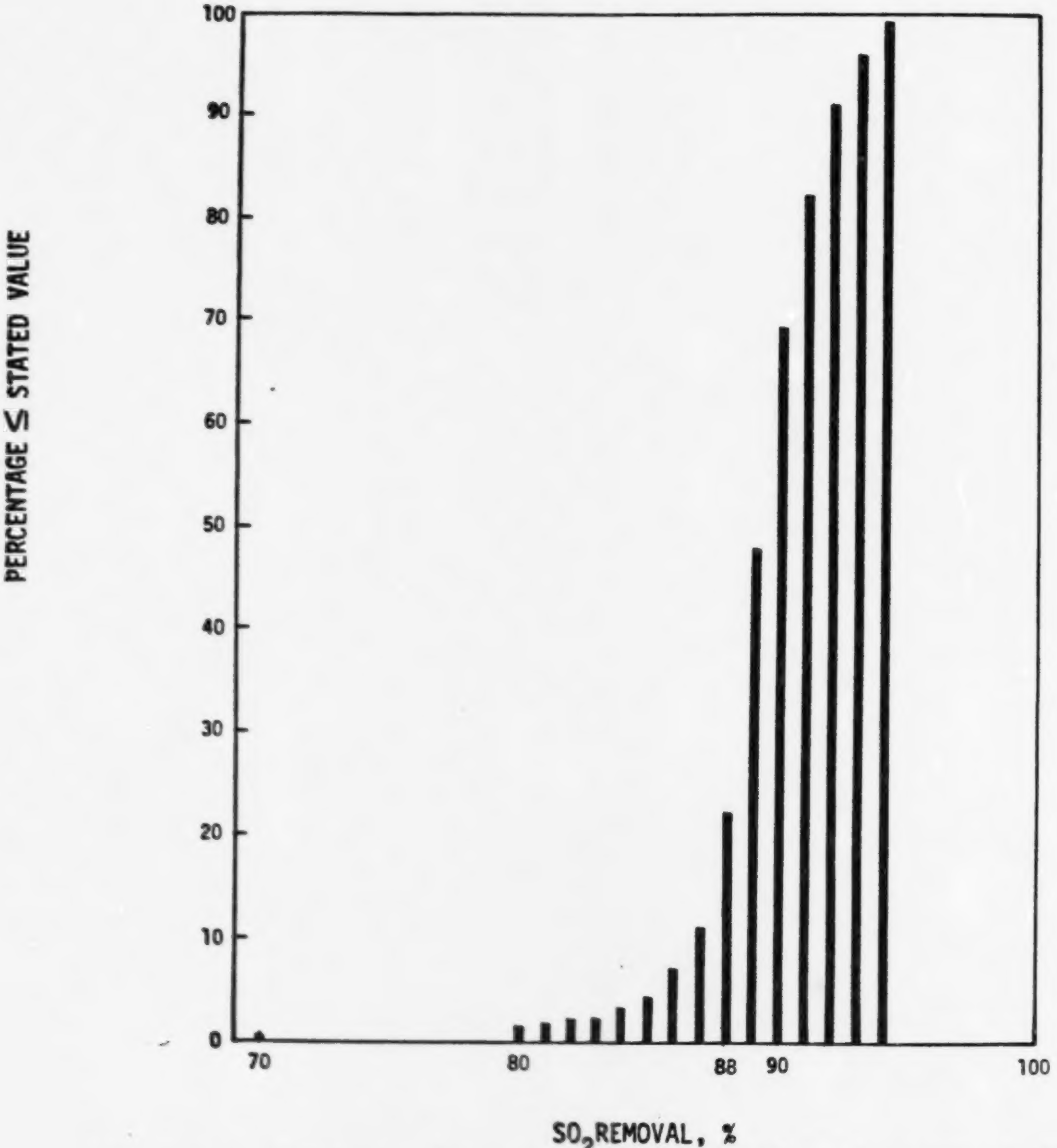


Figure 3-4. Cumulative frequency distribution of one-hour average SO_2 removal.

3. September 25 (79%) - solution feed rates were too low to maintain removal while building feed tank level.

The absorber was operated to achieve about an 89% or higher removal because operating control for SO_2 removal was set to reduce the SO_2 concentration on a diluted basis by 90%. Because dilution amounted to about 10%, primarily as moisture, the actual removal was about 1% less than the reduction in concentration. The SO_2 removal data presented in this report has been corrected for dilution. Removals of 90% and higher are achievable. However, each additional increment of SO_2 removal incurs a penalty in higher evaporator duty and in higher soda ash makeup.

Performance during the second demonstration year was noticeably better than during the first demonstration year (Table 3-2).

TABLE 3-2. COMPARISON OF SO_2 REMOVAL FOR FIRST AND SECOND DEMONSTRATION YEAR

	Percent of time (1-hour average)	
Removal	First year	Second year
90% and greater	45	52
89% and greater	66	78

The SO_2 removed per boiler heat input varied from 110 ng/J to 400 ng/J (0.25 to 0.94 lb/million Btu), well within the Federal new source performance standard of 520 ng/J (1.20 lb/million Btu). These results are based on 24-hour data collected during the April-September period (Table 3-2A). The reported data are limited to those days that valid 24-hour averages during Mode 2 operation were available.

TABLE 3-2A. SO_2 EMISSIONS PER BOILER HEAT INPUT

Month	No. of Days	SO_2 Emitted, ng/J ($lb/10^6$ Btu) Range		Average	
April	14	310-400	(0.73-0.94)	370	(0.85)
May	6	250-280	(0.58-0.66)	270	(0.62)
June	6	110-240	(0.25-0.55)	190	(0.45)
July	6	210-280	(0.48-0.65)	250	(0.59)
Sept	6	220-370	(0.51-0.85)	300	(0.69)

Feed Characteristics

Because removal efficiencies were in the range of 85 to 93% for more than 95% of the time, we have assumed this to be the steady state condition of the process. Thus, the process has been proved operable within the variabilities of feed conditions that were experienced.

Sulfur level in the as received coal averaged 3.09 wt%. Confidence limits for this mean value at the 95% level were 2.99 to 3.19 wt%. A coal of 3.16 wt% sulfur was used for design of the FGD plant and is within these confidence limits, confirming that the actual sulfur level in the coal did meet design conditions. The estimated range, based on a distribution of two standard deviations from the mean, was 2.6 to 3.6 wt%. This was a slightly wider range than the expected limitations of about 2.8 to 3.5%.

Concentration of SO_2 was expected to be 2,185 ppmv (wet basis) at a coal sulfur level of 3.16 wt%. The equivalent SO_2 feed rate for this concentration is 0.6101 kg/s (4,842 lb/h) at the design flue gas flow rate of 151 m^3/s (320,000 acfm) and the design equivalent boiler load factor of 80%. Actual flue gas flows were somewhat higher due to higher than expected excess air caused by excessive inleakage (believed to be at the air preheaters). The absorber was designed to receive a volume of flue gas equivalent to a 100% load factor and to remove the corresponding amount of SO_2 at inlet concentrations of at least 2,185 ppmv. While the capacity to handle flue gas flow rates and SO_2 feed rates equivalent to a 100% load factor was demonstrated for short periods, sustained operation was possible only at load factors slightly better than 80% due to the limited capacity of the evaporator and limited surge capacity in the regeneration loop, as designed. This limited the capacity of the boiler during high demand and was in addition to the derating effect from FGD plant steam consumption. While performance at high load was not fully demonstrable, minimum sustainable operating rates were demonstrated during turndown tests. Full operation at 53 MWe was sustained for four days. Table 3-3 summarizes the inlet conditions of this test.

Below this operating level, the FGD plant is limited by the turndown capability of the reduction unit. The absorber/evaporator was operated down to 44 MWe. It must be remembered, however, that these minimum loads are generator output after derating due to the steam requirements of the FGD plant. Equivalent load relative to flue gas flow and boiler heat input are 50 MW at the 44 MW level and 60 MW at the 53 MW level. In subsequent discussions, the "equivalent load" will be called the generating potential of the boiler. The feed and boiler operating characteristics of SO_2 feed, flue gas flow rate, excess air, boiler heat input, and boiler load as they affect SO_2 removal are discussed below.

TABLE 3-3. FGD PLANT MINIMUM SUSTAINABLE LOAD TEST RESULTS

Parameter	Unit
Length of test	4 days
Average load	53 MWe
Flue gas flow, m^3/s (acfm)	112 (237,000)
SO_2 feed, kg/s (lb/h)	0.426 (3,381)
SO_2 feed	2,069 ppmv wet
Oxygen in flue gas	7.5 vol.% dry
SO_2 removal	88-93%

Sulfur Dioxide Feed--

Thirty-day average feed rates and the maximum feed rate that could be sustained for at least 24 hours are shown for each month in Table 3-4. The sustained feed rates represent the maximum capacity of the FGD plant demonstrated during these tests. Higher feed rates, without a loss of SO_2 removal efficiency, are limited by the design capacity of 80% of full load of the recovery unit. Figure 3-5 shows that the maximum design feed rate could be met and exceeded.

The ranges of estimated error of the SO_2 feed rates are shown in Figure 3-5. The SO_2 feed rates were determined either from measured concentrations and calculated flue gas flow rates or from sulfur content

TABLE 3-4. THIRTY DAY AVERAGE AND SUSTAINABLE SO$_2$ FEED RATES

Month	30-day average		Sustained[*]	
	kg/s	lb/h	kg/s	lb/h
October 78	0.654	5,190	0.683	5,420
November	0.552	4,380	0.636	5,050
December	0.558	4,430	0.605	4.800
January 79[†]				
February[†]				
March	0.596	4,730	0.615	4,880
April	0.617	4,900	0.687	5,450
May	0.638	5,060	0.692	5,490
June	0.517	4,100	0.631	5,010
July	0.556	4,410	0.750	5,950
August	0.554	4,400	0.612	4,860
September	0.629	4,990	0.732	5,810
October	0.573	4,550	0.620	4,920

[*] Sulfur dioxide maximum feed rate that was sustained for at least one day.

[†] Not determined - insufficient data.

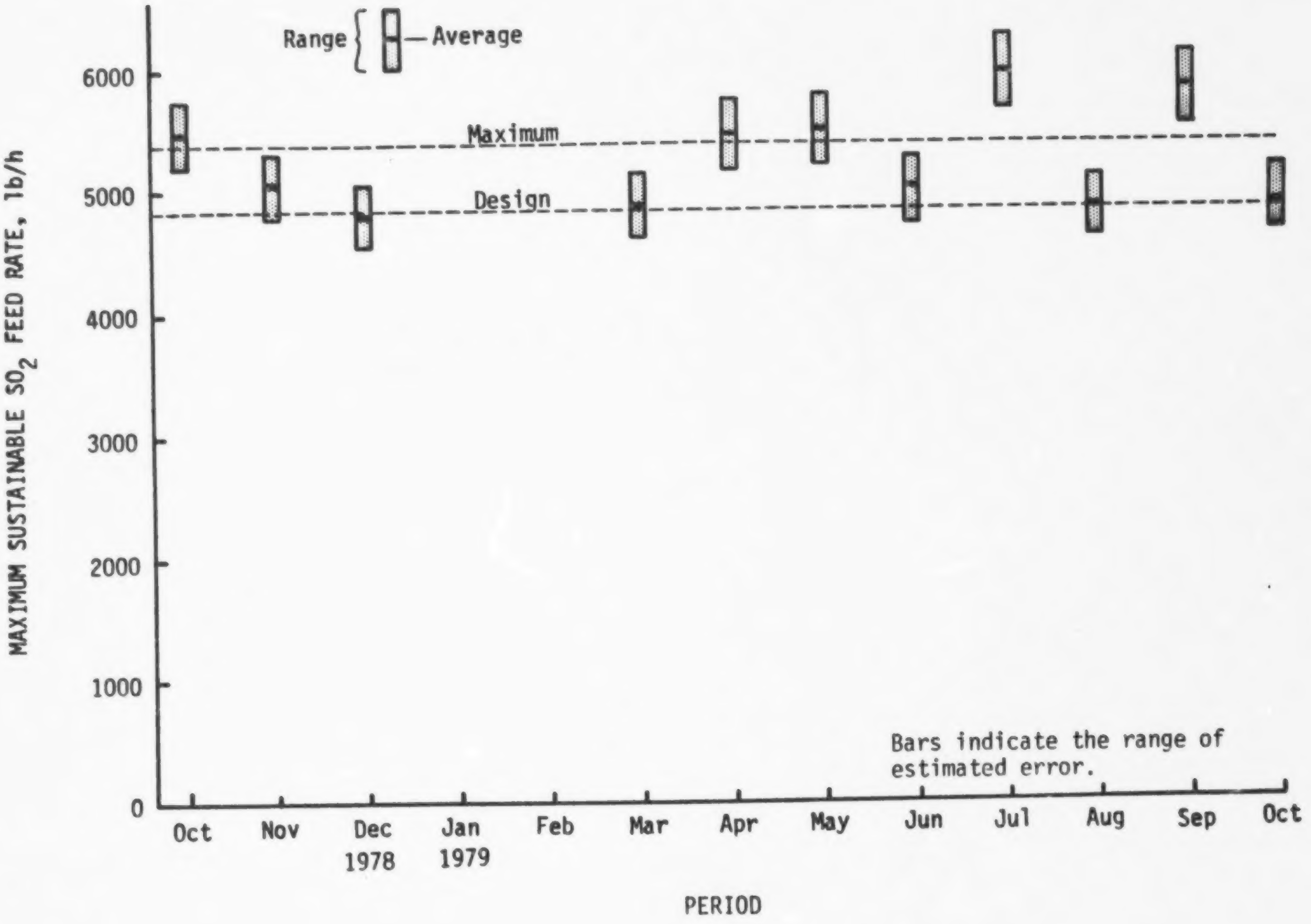


Figure 3-5. Maximum sustainable SO_2 feed rates.

of the coal and measured coal rates. If flue gas flow data was unavailable or limited, the coal data was used with the assumption that about 98% of the sulfur ends up in the flue gas. Sulfur balance determinations across the FGD plant show that about 15% of the inlet sulfur is unaccounted for. This is assumed to be due to a combination of unmeasured losses and of errors of measurement or estimation. The uncertainty in the SO_2 feed rates is estimated at ±8% when using the data obtained from flue gas measurements.

Boiler Load--

SO_2 removal rates of 89% or better were attained at 24-hour average loads from 53 MWe to 93 MWe. However, 85 MWe was usually the upper limit of operation. Figure 3-6 presents the load factors for each 30-day period of the second demonstration year during Mode 2 operation of the FGD plant. The FGD plant was expected to operate at a boiler generating potential of 92 MW and did, indeed, meet or exceed this capacity during 3 of the 30-day periods. This is the generating potential the boiler would be expected to achieve had it not been derated due to steam consumption by the FGD plant. For the rest of the time the boiler was demand limited and the load factors remained below the 92 MW level.

Flue Gas Flow Rate

While boiler load was generally below the design level, the reverse was shown for flue gas flows. This was due to higher than expected excess air in the flue gas. Figure 3-7 shows flue gas flow rates as a function of both the actual generator output and the boiler generating potential. Flue gas flows of 151 m^3/s (320,000 acfm), the design level, were attained at 71 MW of generator output or 80 MW of generating

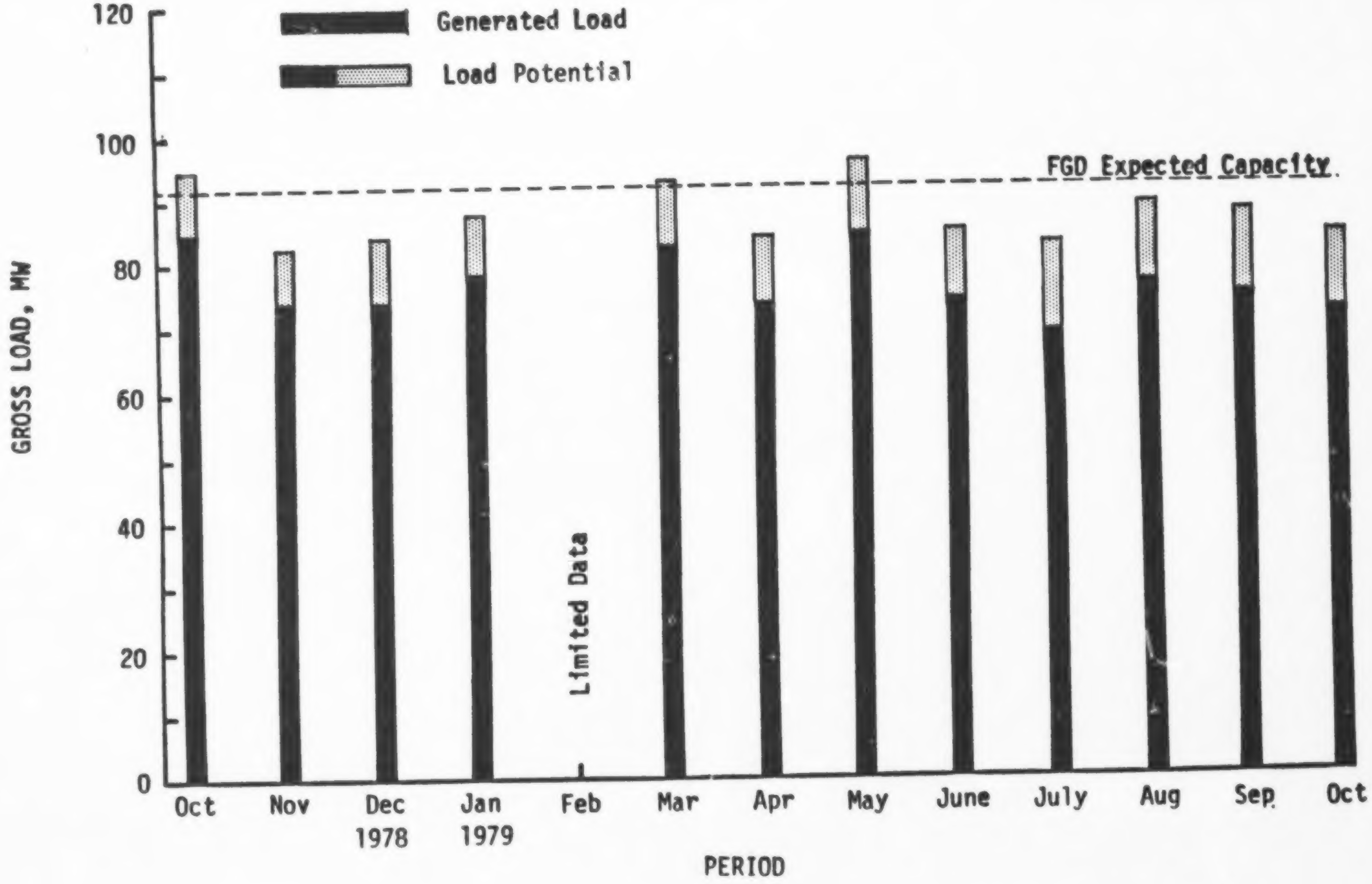


Figure 3-6. Boiler load and load loss during FGD operation.

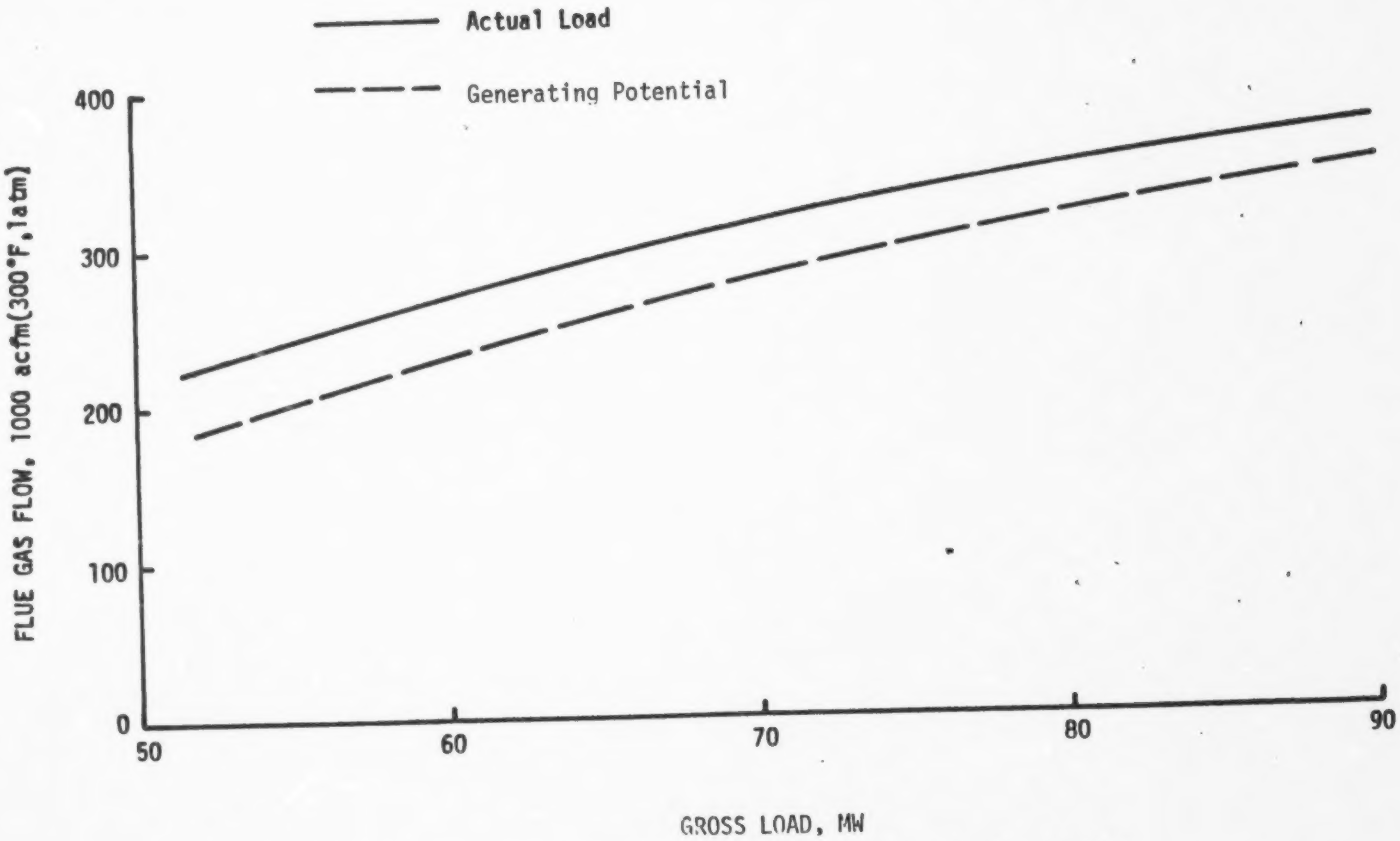


Figure 3-7. Flue gas flow rates at actual load and load potential.

potential. At a FGD load limit of 85 MW, the flue gas flow rate was over 170 m^3/s (360,000 acfm). At a generating potential of 92 MW, the design point, the flue gas flow rate was nearly the same (360,000 acfm). Because the absorber was designed for full load, the greater volume of flue gas presented no apparent problem. A sulfur dioxide removal performance of 89% was attained at flue gas flows within the limits of 112 m^3/s (237,000 acfm), at minimum sustainable load, and 180 m^3/s (380,000 acfm). However, the more typical flow rates varied only from 149 m^3/s (315,000 acfm) to 172 m^3/s (365,000 acfm).

The high flue gas flow rates are explained by the higher than expected amount of excess air which can be attributed primarily to inleakage air, believed to be entering at the air preheaters. Oxygen levels in the flue gas averaged 8.0% by volume compared to an expected oxygen level of 5.6%. These oxygen levels vary with load because the amount of excess air for combustion varies inversely with load below about 92 MW. Typically, the additional amount of air would add about 17% to the quantity of flue gas. At higher than normal loads, the excess air averaged slightly less than the overall average and would add slightly less than 17% to the quantity of flue gas.

Inlet Concentration of SO_2

Average and minimum/maximum concentrations are shown in Table 3-5. Typical data groupings were 200 ppm above and below the 30-day average resulting in a typical range of values from 1,700 ppm to 2,400 ppm. In June 1979, the minimum SO_2 concentration of 598 ppm was the last hourly average before the reduction unit was shut down because of insufficient feed to operate. The low SO_2 concentration resulted from feeding a

TABLE 3-5. INLET CONCENTRATION OF SO_2

Month	Sulfur dioxide in ppmv, wet[*]		
	Average	Minimum	Maximum
October 78	2,148	1,290	2,293
November	2,239	2,100	2,500
December	2,174	1,680	2,437
January 79	2,180	1,210	2,417
February[†]	-	-	-
March	2,155	1,781	2,291
April	2,045	1,150	2,449
May	2,203	1,609	2,771
June	1,873	598	2,278
July	2,087	1,664	2,521
August	1,801	1,625	2,078
September	2,113	1,469	2,688
October 79	2,148	1,620	2,437
All	2,097	598	2,771

[*] One-hour averages.

[†] Limited data.

blend of a medium sulfur and Captain coal. The sulfur content of the medium sulfur coal was slightly higher than 1 wt.%. A gradual decrease in inlet sulfur occurred over a three-day period before shutdown was required. Sulfur dioxide removal was unaffected. Pertinent data for the last 24 hours before shutdown were as follows:

- Average inlet SO_2 concentration, ppmv: 1,163
- Average SO_2 feed rate kg/s (lb/h): 0.2960 (2,349)
- Average SO_2 removal, %: 91

FGD PLANT DEPENDABILITY

The various modes of operation, depending on the availability of equipment and on flue gas availability, are restated as follows:

Mode 1 - Boiler operating, FGD not operating.

Mode 2 - Integrated operation of the absorber/evaporator and reduction units without bypass of any of the flue gas.

Mode 3 - Operation of absorber/evaporator with the flue gas bypass damper open. Bypassing of the flue gas may or may not occur depending on the booster blower speed setting.

Mode 4 - Operation of absorber/evaporator with the reduction unit not operating. Recovered SO_2 (from evaporator overhead) is vented after dilution with flue gas from other units.

The FGD unit was considered to be fully operable only during Mode 2, although that also may have been the case during part of the Mode 3 operation. The exact operating status was difficult to determine while the bypass damper was open. However, failure to include any part of Mode 3 operation as fully operable time did not penalize the FGD process unfairly because the amount of accumulated full operation time with the bypass damper open was very low. Mode 2, or full operation status, does not consider the operation or performance of the purge treatment unit. The purge unit may or may not have been operating during Mode 2 operation and problems with the purge unit are discussed later (see Design Limitations). Subsets of operating Modes 3 and 4 are listed below:

Mode 3A - Bypass open, outward flow. The SO_2 removal data is valid at uncertain feed rates.

Mode 3B - Bypass open, inward flow of air or of flue gas from Boiler No. 6. The SO_2 removal data is invalid because the inward flowing gas does not flow past the SO_2 sampling position.

Mode 4A - Reduction unit down, bypass shut. SO_2 removal data is valid at known feed rates.

Mode 4B - Reduction unit down, bypass open. SO_2 removal data may or may not be valid depending on inward or outward flow.

During Mode 3, low static pressure in the duct between the ID fan and the booster fan indicates inward flow of undetermined magnitude whereas high static pressure indicates outward flow. For data analysis purposes, data during Modes 3B and 4B were ignored and data of Modes 3A and 4A were included in the data base, if the inlet static pressure was normal.

Dependability encompasses a range of indices that define the amount of online time of the FGD plant or the boiler. The viability parameters were used as defined in the EPA Utility FGD Survey[10] (Table 3-6). The definition of "called upon" hours for determining the reliability index follows.

TABLE 3-6. DEFINITION OF VIABILITY INDICES*

Parameter	Description
Boiler Capacity Factor	(kWh generation in year)/(maximum continous generating capacity in kW x 8,760 h/yr).
Boiler utilization parameter	Hours boiler operated/hours in period, expressed as a percentage.
Efficiency - particulates	Operational - The actual percentage of particulates removed by the FGD system and the particulate control devices from the untreated flue gas. All others - The design efficiency (%) of particulate removed by the FGD system and the particulate control devices.
Sulfur Dioxide	Operational - The actual percentage of SO_2 removed from the flue gas. All others - The design efficiency.
FGD availability factor	Hours the FGD system was available for operation (whether operated or not)/ hours in period, expressed as a percentage.
FGD reliability factor	Hours the FGD system operated/hours FGD system was called upon to operate, expressed as a percentage.
FGD operability factor	Hours the FGD system operated/boiler operating hours in period, expressed as a percentage.
FGD utilization factor	Hours the FGD system operated/hours in period, expressed as a percentage.
FGD status - Category 1	Operational - Unit has been or is in service removing SO_2.
- Category 2	Under construction - Ground has been broken for installation of FGD system that has not become operational.
- Category 3	Planned, contract awarded - Contract has been signed for purchase of FGD system but ground has not been broken for installation.

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*Melia, M., M. Smith, T. Koger, and B. Laseke. EPA Utility FGD Survey. Contract No. 68-02-2603, U.S. Environmental Protection Agency, February 1979.

Certain design decisions were made which limited the ability of the FGD plant to follow the full range of normal boiler operation (operability). Nonetheless, design changes could be made or redundancy provided on another installation that would maximize the FGD unit's ability to follow boiler operation. In this report, dependability is assessed relative to the specific design features of this FGD unit. Accordingly, the reliability of the FGD plant is defined as its ability to follow boiler operation only when specific criteria are met. Thus, the FGD plant reliability is determined only for those hours that it is "called upon" to operate, due to essential feed streams being available simultaneously. These include:

1. Flue gas at rates not less than 46 MW equivalent
2. Stable steam pressures within an operable range around a design pressure of 3,790 kPa (550 psig).
3. Electricity
4. Natural gas
5. Soda ash
6. Boiler stable within limits of greater than 46 gross MWe, and coal sulfur content greater than 2.8% and less than 3.5%.

Also, the FGD plant cannot operate for sustained periods at flue gas rates above the equivalent of 85 MWe.

The significant viability indices for this demonstration program are reliability, operability, and utilization. These indices are described below:

- Reliability - hours FGD operated/hours called upon to operate. This parameter defines the dependability of the FGD unit relative to its design capability.
- Operability - hours FGD operated/hours boiler operated. This parameter defines pollutant reduction performance dependability, or the time that SO_2 emissions from the boiler are being controlled. It also is useful for examining partial operation and the dependability of individual subsystems.
- Utilization - hours FGD operated/hours in period. Low utilization affects operating costs significantly because fixed charges, labor costs, and overheads continue to accrue while the FGD unit is inactive.

Detailed operating records for determining the viability factors are included in Appendix A.

Reliability

Figure 3-8 shows the reliability (hours of Mode 2 operation/hours called upon to operate) for the 13 months of the second demonstration year. Called upon hours are those boiler operating hours when the boiler is delivering flue gas and steam within the design range. Figure 3-8 shows the FGD unit reliability factors plotted for each of the 13 months. The overall average reliability was 61%. The reliability record was established with virtually no redundancy built into the FGD unit. Also, the evaporator was designed for only 80% of full boiler load. Due to limited surge capacity within the regeneration loop, the FGD plant was unable to operate in a full operation mode during evaporator or reduction

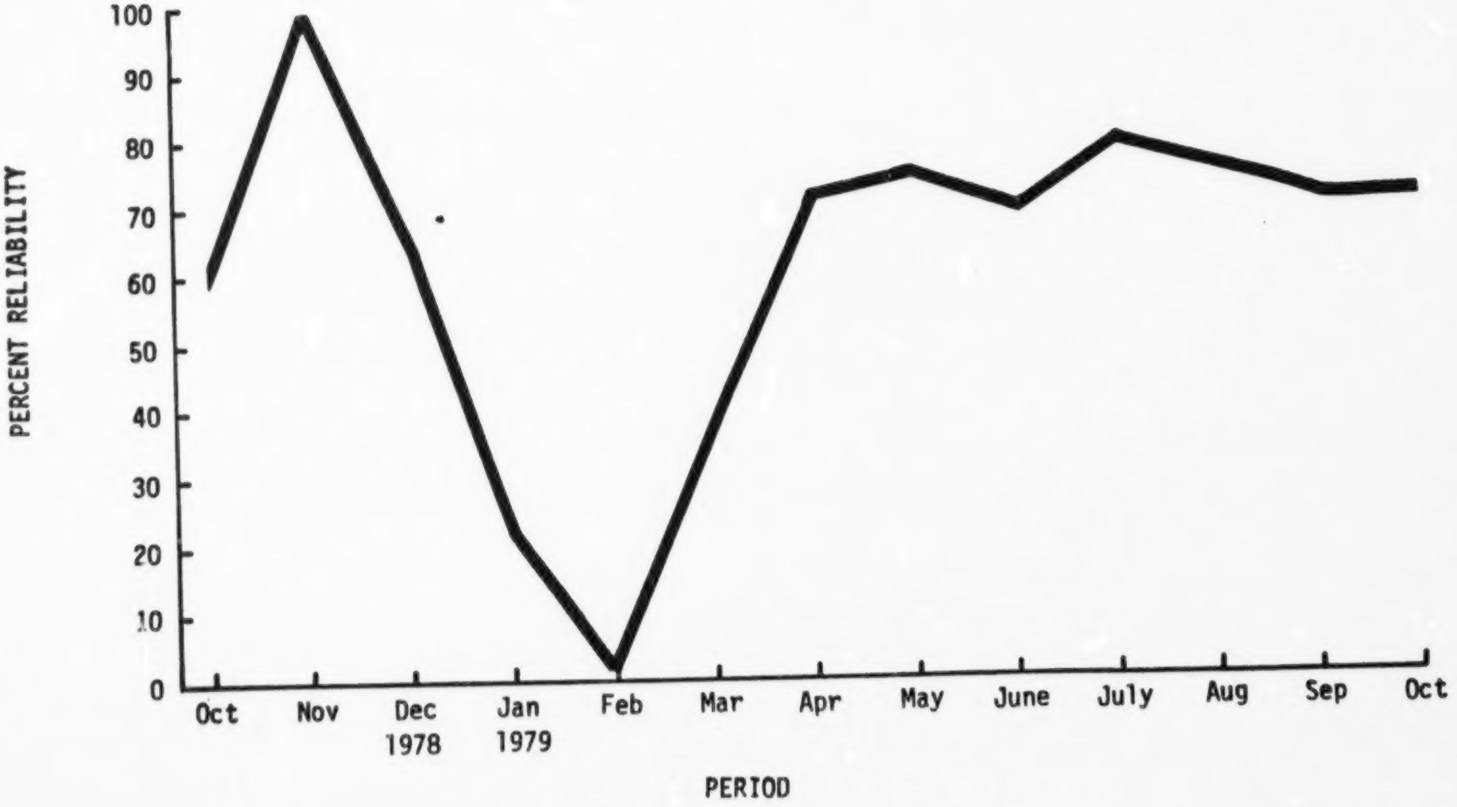


Figure 3-8. FGD reliability index.

unit shutdowns. The ups and downs of operating performance shown on Figure 3-8 may be summarized as follows:

- Best reliability was achieved during October and November 1978. For a 57-day period, October 16 through December 11, interruptions were minor and shortlived. FGD reliability was 99% during November.
- Operation was limited to 66% reliability during December to clean the evaporator heater and for reduction area repairs.
- The FGD plant went down on January 10 for 43 days to repack the evaporator circulating pump and to retube a sulfur condenser.
- Full operation was limited during late February and early March due to numerous mechanical problems and leaks. The major problem during March was an outage of 16 days to overhaul and realign the evaporator circulating pump.
- From April until October, FGD plant reliability averaged 73%. Recurring problems with the booster fan turbine speed control and with the reduction unit were the primary limitations to better reliability.

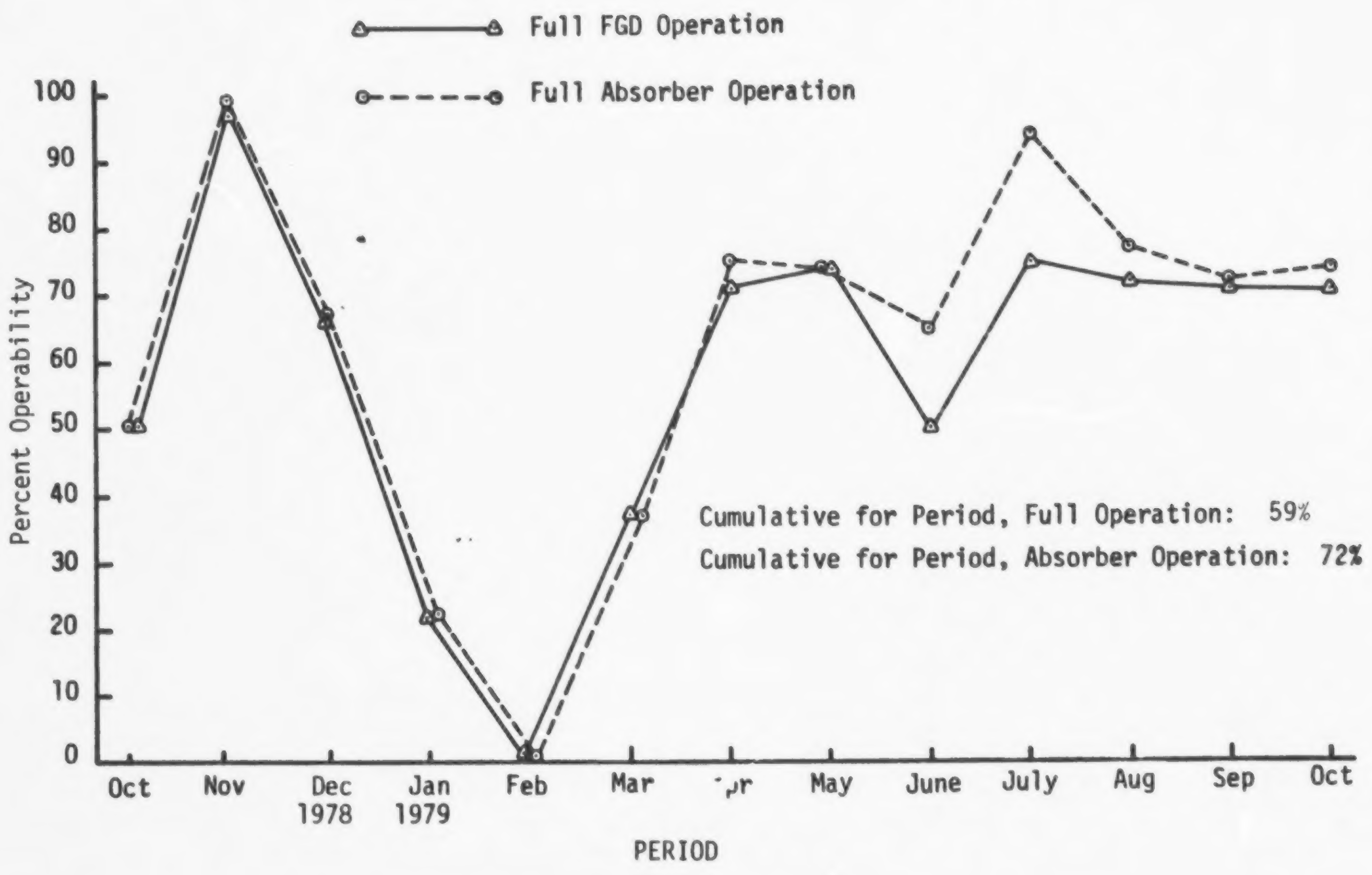
It should be noted that during the 7-month period, April to October, month-by-month reliabilities were fairly consistent, were primarily in a range of 70-75%, and were 13% higher than the 13-month average reliability.

Operability

Operability at full operation was 59%, only 2% less than the reliability. The majority of the difference is attributable to inlet sulfur below design as a result of burning medium sulfur coal. The operability factor has been adjusted for no operation during baseline testing in October 1978 and for partial operation during special testing in July 1979.

Figure 3-9 compares operability at full operation with operability during absorber/evaporator operation. The absorber and the SO_2 recovery unit were significantly more dependable than the FGD plant as a whole. Included are the booster fan, prescrubber, absorber, and evaporator but not the purge treatment and reduction units. The absorber/evaporator could continue to operate while the reduction unit was down by venting the recovered SO_2 to the Unit No. 6 stack. Figure 3-9 shows that the absorber was operated for 72% of the boiler operating time. For the period April to October, total absorber operating time was 87% of the boiler operating time. The additional absorber/evaporator operating time usually occurred during reduction unit outages, operation with bypass open, and during startups.

Operation with the bypass damper open was infrequent and usually of short duration. Operation on bypass occurred during only 4% of boiler operation (Figure 3-10). To protect the boiler, controls were provided to trip the damper open on high or low static pressure in the duct between the ID fan and the booster fan. For example, if the booster fan tripped off, pressure would build and trip the damper open. Normal operating procedure was to open the bypass damper if the booster fan was on manual control. The damper tripped open on high pressure on only three occasions. Most of the time that the bypass damper was open may be attributed to startups, to mechanical and control problems of the damper itself, and as a precaution during repairs to or unstable



Cumulative for Period, Full Operation: 59%
Cumulative for Period, Absorber Operation: 72%

Figure 3-9. FGD plant operating history.

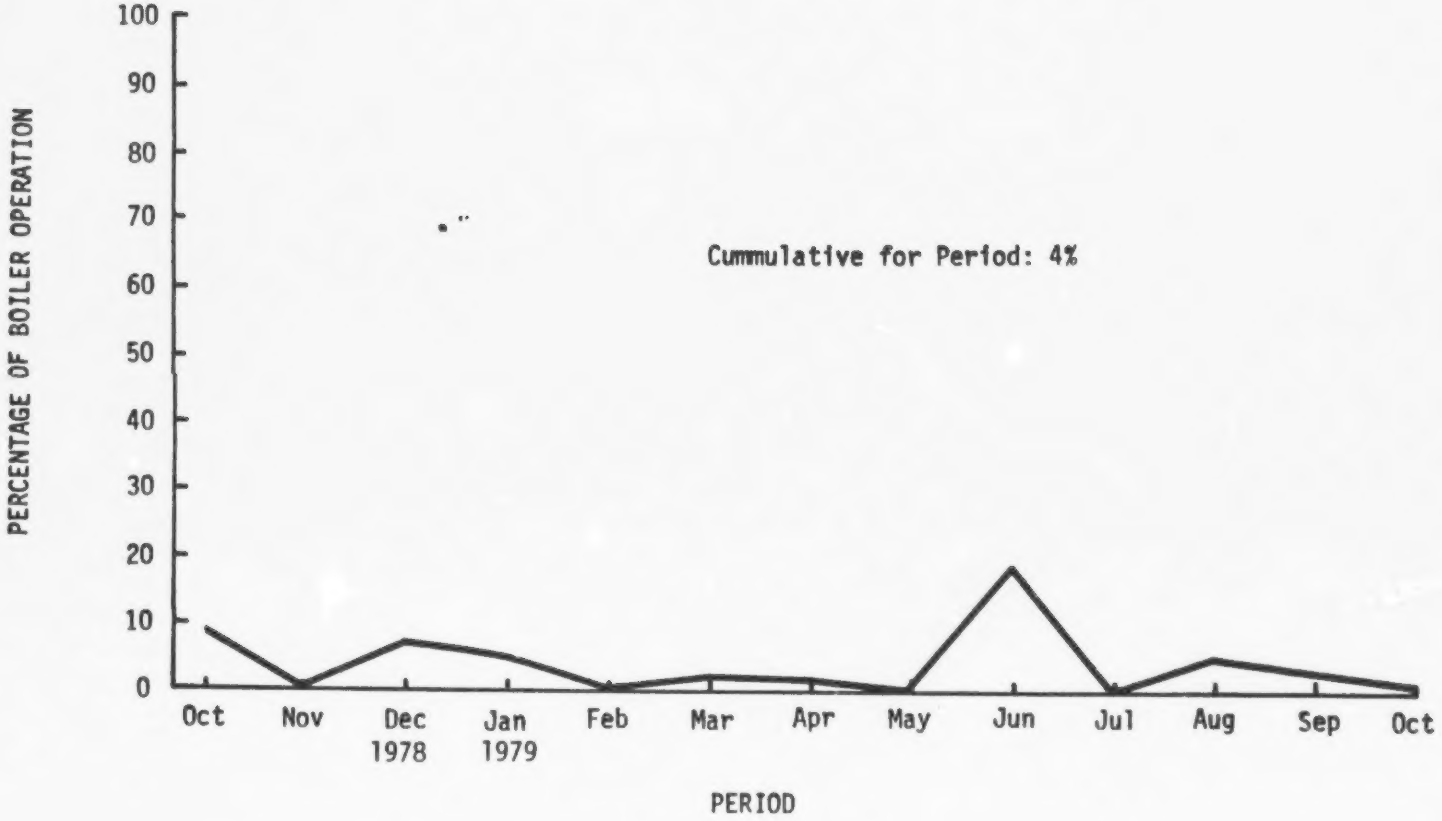


Figure 3-10. FGD full operation with bypass damper open.

operation of other components of the system. The FGD unit was on bypass for 111 hours in June as a precaution during the time that the west ID fan was speed limited due to a bearing problem. A complete record of operation on bypass is given in Appendix A.

Utilization

The FGD plant was utilized during 56% of the hours in the period, including adjustments for baseline and special testing. This amounts to 168 days of partial or no operation during which substantial costs continued to accrue. FGD plant interruptions were due to five factors:

1. Boiler down
2. Boiler operating out of FGD design limits
3. Down or at partial operation for testing
4. Forced shutdown or forced partial operation for repairs
5. Startup and shutdown time

There were 9,369 hours in the thirteen 30-day periods from October 3, 1978 to October 29, 1979. Baseline and special testing used 224 hours of the total, and these hours were not counted in determining dependability factors. The effect on utilization of the remaining four factors is displayed on the bar graph of Figure 3-11.

REASONS FOR INTERRUPTIONS

Table 3-7 identifies the equipment items that gave the most problems. The highest percentage of downtime was due to numerous interruptions in SO_2 reduction unit operation. Because surge capacity for the scrubbing solution was minimal, any interruption of reduction unit operation required that either the evaporator be shut down or recovered SO_2 be vented. The usual practice was to vent the SO_2 to the Unit No. 6 stack.

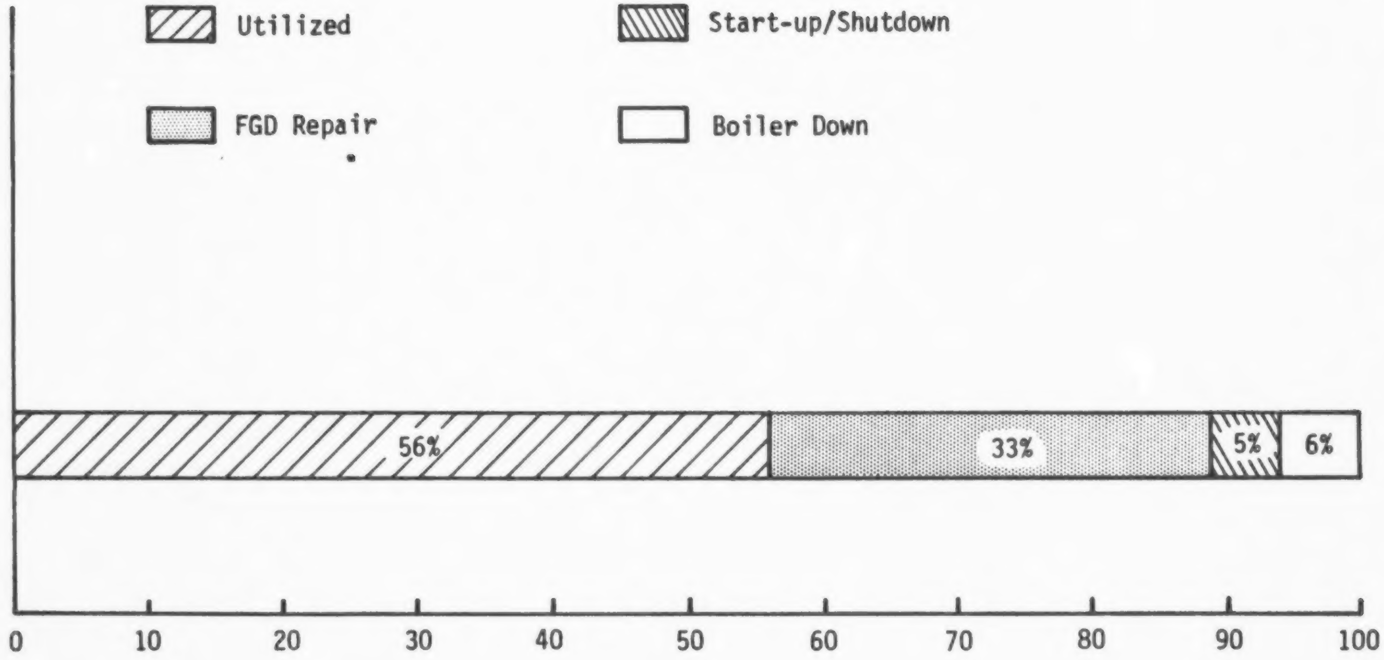


Figure 3-11. FGD plant limits to utilization.

TABLE 3-7. REASONS FOR INTERRUPTION TO OPERATIONS

Equipment or reasons	Days of interruption	% of called upon time
Reduction unit	59	17
Evaporator circulating pump	28	8
Booster blower	18	5
Startup and shutdown	17	5
Other equipment, including absorber	10	3
Evaporator	8	2

The major interruption of the reduction unit was a 35-day shutdown to retube a sulfur condenser. Without this interruption, the reduction unit prevented full operation during 7% rather than 17% of the called upon hours. The most frequent outages were those due to sulfur deposition, leaks, and valve repairs.

With electric instead of steam turbine drive for the evaporator circulating pump, emergency shutdown of the FGD plant was accomplished several times without difficulty. There were eight boiler shutdowns while the FGD plant was at full operation. Four of the shutdowns were without warning; two of the shutdowns were after a short warning period of less than one hour; and the other two occurred with adequate advance notice, and the FGD shutdowns preceded the boiler shutdowns by six hours and nine hours.

Startups were a different story. Table 3-8 shows the startup time required for the same eight boiler shutdowns. The usual sequence after boiler startup was to start the absorber/evaporator loop and flue gas flow first, followed by the reduction unit, after which the bypass damper was closed. The startup record of Table 3-8 indicates, perhaps, that more time is required for startup after the more lengthy shutdowns. Otherwise, some of the startups seem to be unnecessarily long.

TABLE 3-8. FGD PLANT STARTUP TIME

Boiler down h	FGD start-up time h	Startup sequence, h		
		Absorber/evaporator	Reduction	Bypass Damper
3	1	0	0.3	0.7
3	0.2	0	0.2	0
34	31	9	22	0.2
100	84	8	47	29
3	38	3	31	4
138	70	24	21	25
11	11	2	4	5
3	4	0	3	1
295	239	46	129	65

Other reasons for FGD plant interruptions may be summarized as follows:

- The evaporator circulating pump was down three times for repacking, for overhaul, and for realignment of the motor shaft. It also was down once to replace a seal attributable to interruption of the steam supply from the boiler. Without steam, the condensate used for seal water was lost.
- The booster blower was down for relatively short periods but frequently. Most of the problems centered around the turbine governor and the gear reducer. There were no problems associated with the internal surfaces of the fan itself.
- There was one interruption for cleaning the evaporator heater after it had plugged.

- The absorber operated essentially trouble free. There was only one 6-hour interruption caused by an obstruction in the process water valve.
- Other problems accounted for less than 3% of the called upon time. They include frequent replacement of the SO_2 superheater with an overhauled spare and repairs to the SO_2 compressor. These items of equipment are in the line feeding SO_2 to the reduction unit. Other interruptions were for instrument and duct leak repairs.

A more detailed description of FGD plant interruptions and when they occurred is included in Appendix A.

DESIGN LIMITATIONS

This WL/Allied unit was designed for the specific flue gas rates and flue gas characteristics expected from the Mitchell No. 11 boiler. Furthermore, virtually no redundancy was built into the FGD unit and the evaporator was designed to regenerate only the SO_2 removed that was equivalent to the expected SO_2 feed rate at 92 MW. The design boiler load was the expected load factor and the evaporator would have had sufficient capacity if sufficient surge capacity for the absorber feed and spent solutions had been provided. With limited surge capacity within the regeneration loop, the FGD plant was unable to operate in a full operation mode at SO_2 inlet rates much greater than 92 MW generating potential. The limit was found to be about 95 MW generating potential. Generating potential is the basis of the comparison with the design capacity of 92 MW.

Although not demonstrated in specific tests, the upper limit for the volume of flue gas that the booster fan could handle and that could be passed through the absorber was in excess of full capacity of the boiler. The upper limit for SO_2 feed could not be defined with testing but an evaluation of higher efficiency scrubbing using Davy's design data has suggested some upper limits (see Section 4). Lower limits of flue gas volume and SO_2 feed were not determined below the generating potential of 51 MW. The expected lower limit for stable boiler operation was a 46 MW generating potential. In the full operation mode, the lower limit was a generating potential of 62 MW because of the turndown limitations of the reduction unit.

An unexpected dryer limitation within the purge treatment unit was also a design limitation. Essentially, all the limitations that were evaluated were design limitations, rather than process limitations. These are doubtless correctable when other units are designed. In this section, test results to determine certain limitations of the FGD plant are reported and analyzed. Testing was done to define:

- Minimum sustainable load in the full operation mode
- Absorber turndown
- Load following capability
- Maximum capacity
- Purge treatment limitations

Minimum Sustainable Load

The results of this test were summarized earlier (Table 3-3). The test level was based on expected lower operating limit reported by Allied of 50 to 55 MWe. Thus, four days operation at a steady load within those limits was requested. Overall, for the four days, the SO_2 removed averaged 0.3818 kg/s (3,030 lb/h) at an inlet SO_2 rate of 4,260 kg/s (3,381 lb/h). The average gross load was 53 MWe and the average generating potential at this load was 62 MW. Table 3-9 lists four-day average results for the most important process variables. During the second and third days of the minimum sustainable load tests, data for confirming flue gas flow, SO_2 in and out, and excess oxygen levels were collected by manual methods. These data were in good agreement with the corresponding continuous data (Table 3-10).

TABLE 3-9. MINIMUM SUSTAINABLE LOAD TEST - FOUR DAY RESULTS

Parameter	Units
Average gross load	53 MW
Average generating potential	62 MW
Average boiler heat input, J/s (Btu/h)	163×10^6 (558×10^6)
Average FGD steam consumption, kg/s (lb/h)	6.3498 (50,396)
Average coal rate, kg/s (lb/h)	6.9837 (55,427)
Average sulfur in coal, wt.%	3.08
Average flue gas flow, m^3/s (acfm)	111.9 (237,100)
Average SO_2 in, ppmv	2,059
Average SO_2 inlet rate, kg/s (lb/h)	0.4260 (3,381)
Average SO_2 removal, %	90%
Average SO_2 removal, kg/s (lb/h)	0.3818 (3,030)

TABLE 3-10. MINIMUM SUSTAINABLE LOAD TEST: COMPARISON OF CONTINUOUS AND MANUAL TEST DATA

Hour from 1 to 96 (four days)	Test No. 1 23-28		Test No. 2 46-50		Test No. 3 53-57	
Gross load, MWe	52.4		53.4		53.3	
Coal rate, kg/s (lb/h)	7.184 (57020)		7.101 (56360)		6.980 (55400)	
Sulfur in coal, wt.%	2.92		3.17		3.14	
	Continuous	Manual	Continuous	Manual	Continuous	Manual
Flue gas flow, m^3/s (acfm)	115 (244,000)	147 (311,000)	115 (243,000)	105 (222,000)	113 (240,000)	102 (216,000)
SO_2 in, ppmv wet	2014	2196[2]	2078	2390[2]	2084	2242[2]
SO_2 inlet rate, kg/s (lb/h)	0.3966 (3147)	0.4280 (3397)	0.4061 (3223)	0.4586 (3640)	0.3988 (3165)	0.4196 (3330)
O_2 in flue gas, vol.% dry	6.89	7.21	7.05	9.27[1]	7.29	7.49
H_2O in flue gas, vol.% dry	(3)	7.75	(3)	6.68	(3)	6.15

[1] Questioned.
[2] Includes SO_2 in tail gas.
[3] Not measured.

Absorber Turndown

Although overall system turndown is limited by the reduction area's minimum operating feed rate, tests of the absorber turndown established a minimum throughput equivalent to a generating potential of 50 MW, or 12 MW less than for minimum sustainable load. The measured SO_2 feed rate seems overstated because it was not substantially less than the minimum sustainable load test results, although the coal feed rate was lower. The EPA Method 6 was used to measure SO_2 concentration because continuous monitor data was unavailable, due to loss of temperature and humidity control in the instrument trailer. Therefore, sulfur in the coal has been used to estimate the SO_2 feed rate. Comparisons between absorber turndown and minimum sustainable load are shown in Table 3-11. The limiting factor was flue gas flow rate because booster fan speed was very near the minimum governor setting for the steam turbine drive. Excess oxygen was higher during the absorber turndown test and offset the lesser volume of air free flue gas at the lower coal feed rate. Thus, the flue gas flow data of Table 3-11 show that minimum turndown for flue gas flow was about 100 m^3/s (220,000 acfm).

Maximum Limitations

The absorber was designed to treat a flue gas volume that was in excess of the equivalent of full boiler load. The absorber design limit was 210 m^3/s (450,000 acfm), or 40% higher than the expected operating level of 151 m^3/s (320,000 acfm). Flow rates at the maximum limit or higher were probably achieved for short periods but no data was taken to define the maximum flow specifically.

TABLE 3-11. COMPARISON OF OPERATING CONDITIONS AT ABSORBER TURNDOWN AND MINIMUM SUSTAINABLE LOAD

Parameter	Minimum sustainable load	Absorber turndown
Gross load, MWe	53	44
Generating potential, MW	62	50
SO_2 in, kg/s (lb/h)	0.4358 (3459)	0.5165 (4099)
Coal rate, kg/s (lb/h)	7.089 (56,260)	6.0941 (48,367)
Sulfur in coal, wt.%	3.15	3.28
Flue gas flow m^3/s (acfm) measured	103.5 (219,400)	112.4 (238,200)
Flue gas flow m^3/s (acfm) calculated	104.5 (221,400)	107.8 (228,400)
O_2 in flue gas, vol.% dry	7.35	8.55
H_2O in flue gas, vol.% dry	6.86	6.70

The recovery unit was designed for a boiler generating potential of 92 MW, or 80% of full load. Without adequate surge capacity, a full operation maximum sustainable load could not but a design maximum could be demonstrated. Thus; a sustained test was run at slightly below the design limited load, confirmed by operation, of 85 MWe. This load was sustained for 65 hours.

The probable maximum sustainable SO_2 feed rate, based on an extrapolation of the test data at 80 MWe to 85 MWe, is around 0.63 kg/s (see Table 3-12); however, SO_2 feed rates in excess of 0.63 kg/s (5,000 lb/h)

TABLE 3-12. MAXIMUM SUSTAINABLE LOAD TEST

Parameters	Test data	Projected to 85 MWe
Length of test, hours	65	-
Gross load, MWe	80	85
Generating potential, MW	89	95
Boiler heat input, 10^6J/s (10^6 Btu/h)	254 (869)	270 (923)
Coal rate, kg/s (lb/h)	10.018 (79,513)	10.645 (84,483)
Sulfur in coal, wt.%	3.20	3.20
Flue gas flow, m^3/s (acfm)	152.6 (323,300)	162.1 (343,500)
SO_2 in, ppmv wet	2,050	2,050
SO_2 in kg/s (lb/h)	0.5882 (4,668)	0.6249 (4,960)
SO_2 removed, %	90	90
SO_2 removed, kg/s (lb/h)	0.5273 (4,185)	0.5625 (4,464)

were frequently observed for short periods. The SO_2 removal efficiency was unaffected by the higher feed rates. As previously stated, the absorber is capable of taking much greater volumes of flue gas than the projection to 85 MWe indicates.

Load Following Capability

A test was designed to identify the transient effects on the FGD plant from an unannounced load change. Data were collected every three minutes. Steady state conditions at steady load were defined just before the load change. The boiler operator was requested to change load at the normal rate, beginning at an agreed upon time. The FGD plant operators were aware that the tests were planned but were not advised of the start times. Tests were done at the decreasing and the increasing load conditions of 80 MWe to 57 MWe and 57 MWe to 80 MWe. Data were collected for a sufficient time following the load change to ensure that steady state had been reestablished.

The normal rate of load change of Mitchell No. 11 is 2 MW/min. The actual test rate was about 1 MW/min. This means that the test was not as demanding as had been planned.

Transient effects on the FGD plant were observed during the load changes. However, the performance parameters generally had returned to the steady state condition by the time the load change was completed or shortly thereafter. The significant observations were as follows:

- A decreasing load change of 23 MW, from 80 MWe to 57 MWe, took 20 minutes for a rate of 1.2 MW/min. An increasing load change of 28 MW, from 52 MWe to 81 MWe, took 32 minutes for a rate of 0.9 MW/min.

- With decreasing load, the instantaneous SO_2 removed/MW decreased but had recovered to steady state by completion of the load change. With increasing load, SO_2 removed/MW peaked during the load change and recovered to steady state 10 minutes after completion of the load change.
- With decreasing load, steam consumption/amount of inlet SO_2 and steam consumption/MW increased and remained at the higher level. Steam data were unavailable during the increasing load test.
- With decreasing load, percent of SO_2 removal began to increase during the load change, continued to increase and remained at the higher level. This may have been by operator choice. With increasing load, percent of SO_2 removal dropped slightly during the load change, recovered right after the load change, and showed some cycling after the load change.
- With decreasing load, excess oxygen peaked during the load change. At the air heater inlet it had recovered to steady state by the end of the load change but at a slightly lower level. However, inleakage air increased and the final steady state oxygen level entering the FGD plant was slightly higher. With increasing load, excess oxygen peaked during the load change. At the air heater inlet, the oxygen level returned to about the initial steady state level. With less inleakage air, the final steady state oxygen level entering the FGD plant was lower.

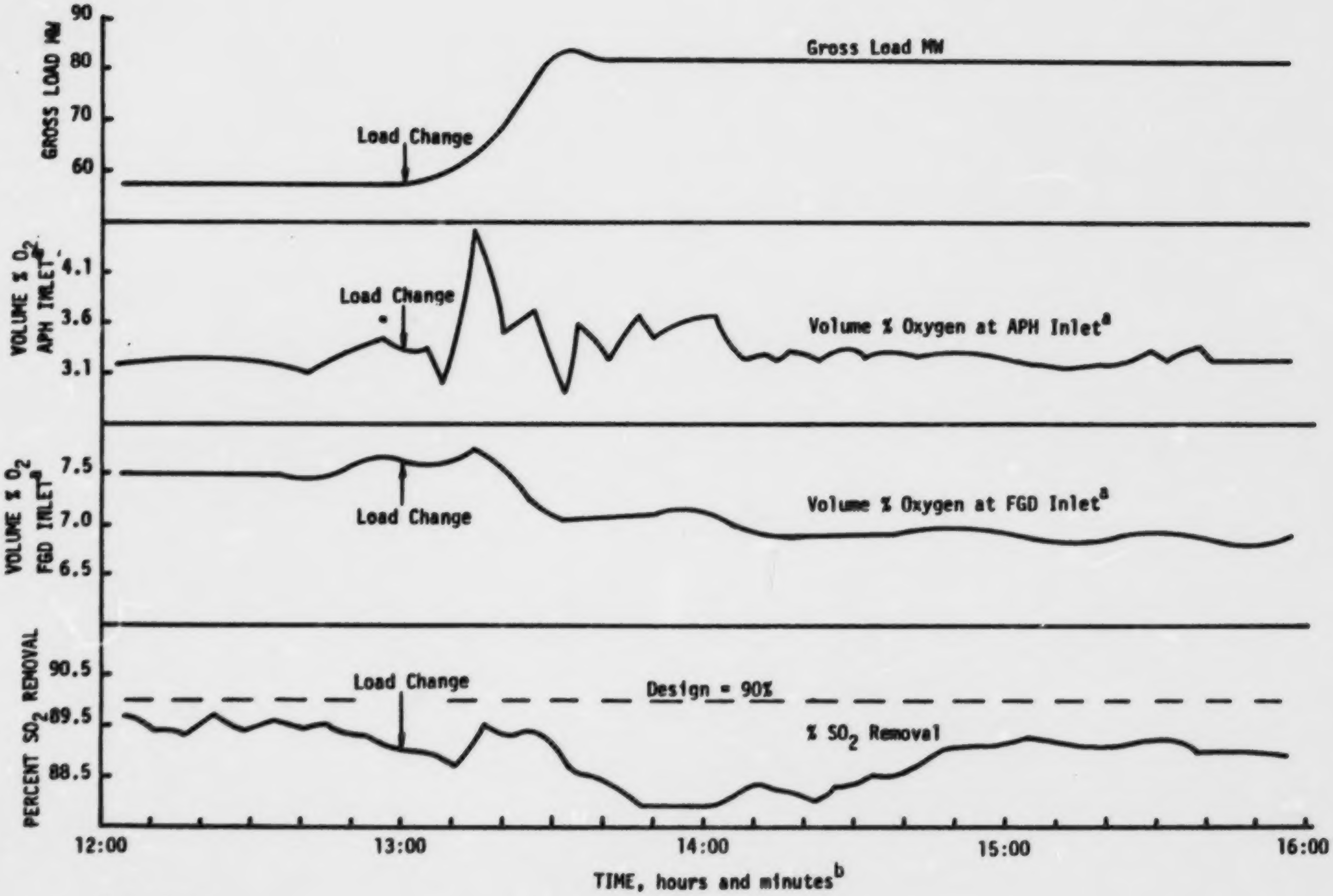
The instantaneous values for the various parameters are plotted with time in Figures 3-12 through 3-15. More of these timelines are included in Appendix A. The variability (standard deviation) of the performance parameters was calculated to compare the load change variability with the steady state variability. Complete data are included in Appendix A.

Purge Treatment Limitations

The purge treatment unit was limited by heat duty underdesign of the purge solids dryer. The purge treatment unit is fed a slipstream of spent absorbing solution. The basic process involves sodium sulfate (Na_2SO_4) crystallization, separation of the crystals from their mother liquor with return of the clear liquor to the evaporator loop, and drying of the crystals to produce a marketable dry product. The "wet" end of this purge treatment system performed satisfactorially, but the dryer had a capacity of only 50% of that needed.

A major objective of regenerable FGD technology is to minimize the size of the purge stream that is usually needed to control the concentration of inert materials in the regeneration loop. For the Wellman-Lord process, this means that the sodium sulfate must be kept at a low level in the absorbing solution. This involves concentration and separation of the sodium sulfate from the absorbing solution, as described above, while minimizing a corresponding removal from the process of active sodium. The sodium values that are purged with the sodium sulfate must be replaced with makeup sodium carbonate.

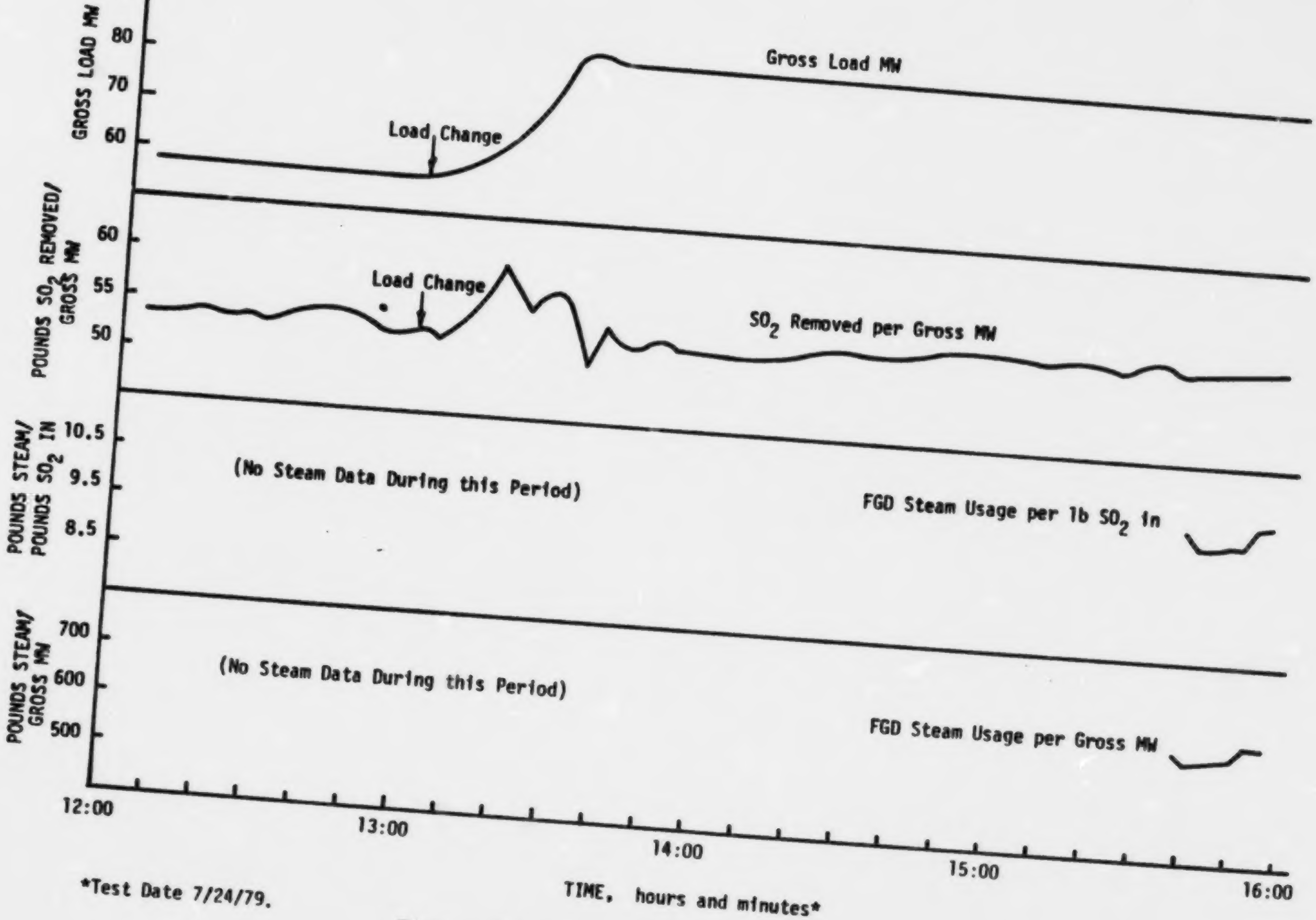
Initially, the test program was to examine the purge treatment system only in relation to raw material (soda ash) consumption and to



[a] Dry Basis.

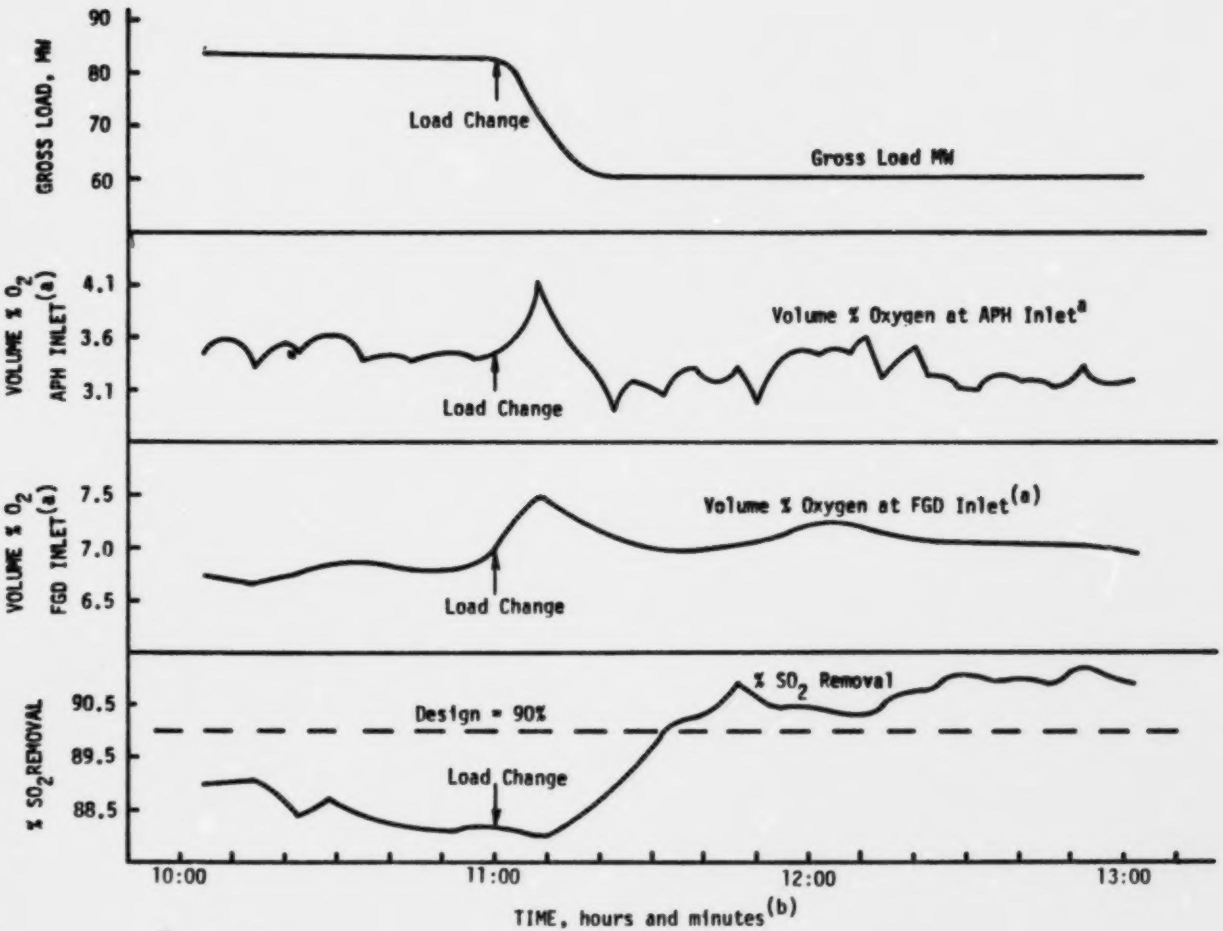
[b] Test Date 7/24/79.

Figure 3-12. 57 to 80 MW load following test.



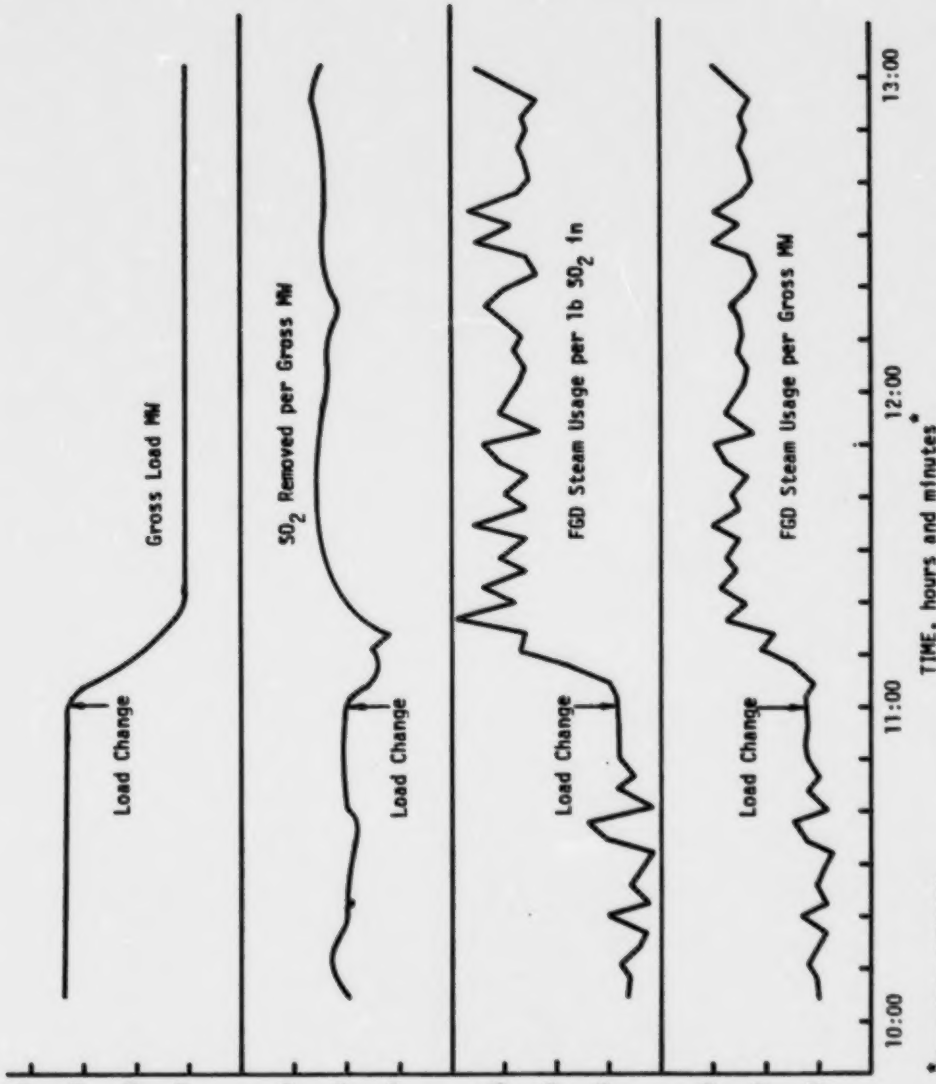
*Test Date 7/24/79.

Figure 3-13. 57 to 80 MW load following test.



[a] Dry Basis
[b] Test Date 7/19/79

Figure 3-14. 80 to 57 MW load following test.



*Test Date 7/19/79

Figure 3-15. 80 to 57 MW load following test.

the byproduct (sodium sulfate) quality. Due to the dryer problem, and because this prompted a need for knowing more about the amount of sulfate ($SO_4^=$) formed in the process, the test contractor provided test support in these areas:

1. Purge rates were calculated from soda ash consumption and the amount of SO_2 removed from the flue gas.
2. Tests were performed in an attempt to determine the amount of sulfate formed in the absorbing solution.
3. Dryer performance tests done by Davy McKee were observed and reported on.

The results of these investigations are as follows.

Purge Rates--

Purge rate is the ratio of moles sodium in purge to moles of SO_2 removed from the flue gas, expressed as a percentage. Purge rates were measured indirectly. They were estimated from the soda ash consumption reported once each month and the calculated hourly average SO_2 removal rates. The purge rate was 21.7% for April through October, the period of the most consistent operation (Table 3-13). The FGD unit was designed for a purge rate of about 10% and may reflect high oxidation rates resulting from the higher than expected excess air levels in the flue gas. Soda ash consumption for the same period averaged 0.091 kg/s (8.7 ton/day), using the cumulative days of absorber/evaporator operation as the time base. The performance guarantee for soda ash consumption was 0.069 kg/s (6.6 ton/day) at the design levels of flue gas flow, inlet SO_2, and excess oxygen.

TABLE 3-13. PURGE RATES*

Month	SO_2 removed		Soda ash consumed,		Purge rate, mole Na/mole S in purge†
	as kg sulfur	(lb sulfur)	kg	(lb)	
April	578,290.4	(1,274,912)	237,900	(524,400)	0.248
May	500,395.9	(1,103,184)	200,800	(442,800)	0.242
June	475,060.9	(1,047,330)	190,900	(420,800)	0.243
July	699,658.1	(1,542,482)	215,300	(474,600)	0.186
August	424,478.6	(935,815)	127,800	(281,800)	0.182
September	547,925.6	(1,207,969)	171,000	(377,000)	0.188
October	467,427.4	(1,030,501)	183,500	(404,600)	0.237
Total period	3,693,282.2	(8,142,293)	1,327,200	(2,926,000)	0.217

*For period, April through October 1979.

†Purge rate determined using soda ash consumption and sulfur removed by absorber.

Sulfate Formation--

The amount of purge to be treated is a function of the formation of sulfate and possibly thiosulfate during absorption and regeneration. Attempts were made to determine the amount of sulfate formation during absorption, but these efforts were frustrated by inability to obtain correct flow measurements and uncertainties about the specific water balance across the absorber. However, the data seem to indicate that sulfate formation is a function of oxygen concentration in the flue gas. The test results are reported in Section 4. Since excess air levels were higher than design expectations, the oxidation rates for sulfate might have been higher than expected, and might require higher than design purge rates. A purge rate of about 10% was the value indicated during the design phase of the project, whereas of actual purge rates up to nearly 25% were estimated. Other than the dryer, there was no indication that the purge treatment unit could not equal or exceed the design requirements.

Purge Dryer Limitation--

Dryer tests performed by Davy McKee determined that the dryer would not work, even at design rates. Since startup, there had always been a question of whether or not the sulfate dryer actually had heat duty design capacity. In May 1979, tests verified that the dryer did not have design capacity. The purpose of the test was to demonstrate the heat capacity of the dryer with a water feed onto a sodium sulfate bed in the dryer. If the maximum rate could be reached with water, the dryer was then to be tested with the sodium sulfate solution recovered from the purge solution.

The maximum dryer capacity achieved during the test was approximately 66% of the design heat duty. Capacities of 59 to 66% of design were maintained for 2-1/2 days. However, after operating for 2-1/2 days at 59 to 66%, the motor tripped out several times because of an amperage overload. When the motor tripped out, a buildup of solids was observed at the discharge end of the dryer. To prevent overloading the motor, the water feed rate was reduced. Thus, during the last 2-1/2 days of the test, the dryer capacity dropped to 45 to 50% of design. At this point, further testing was abandoned. Further details of the tests are given in Section 4.

Possible solutions include:

- A more concentrated feed fed to same dryer.
- A dryer of different design.
- Addition of an antioxidant to the absorbing solution.

The initial design included the feeding of an antioxidant to the absorbing solution to reduce the sulfate formation. Davy McKee has recommended antioxidant treatment for solving the dryer problem, and tests of EDTA (ethylenediaminetetraacetic acid) by Davy McKee were scheduled for early 1980.

ENERGY CONSUMPTION

The FGD plant consumes energy from three sources:

1. Boiler main steam, primarily for process heat for regenerating the spent absorbing solution.
2. Electric power for electric motors, space heating, and lighting.
3. Natural gas, a feed stream for the reduction process and a fuel for incineration of the reduction unit tail gas.

The major requirement was for steam. Table 3-14 summarizes the actual energy provided by the three sources during Mode 2 operation.

TABLE 3-14. AVERAGE HOURLY ENERGY CONSUMPTION

Boiler Source[*]	
Steam	$95{,}309 \times 10^6$ J (90.396×10^6 Btu)
Electricity, metered	$7{,}395 \times 10^6$ J (7.014×10^6 Btu)
Evaporator circulating pump, estimated	$3{,}163 \times 10^6$ J (3.000×10^6 Btu)
Total	$105{,}867 \times 10^6$ J (100.410×10^6 Btu)
Nonboiler source	
Natural gas[†]	$10{,}540 \times 10^6$ J (9.994×10^6 Btu)

[*] Equivalent to the boiler heat input derived from coal.

[†] Used heating value of 38.18 J/m^3 (1025.5 Btu/cf).

The impact of these energy requirements will now be discussed in some detail.

Steam

Steam requirements were expected to be 7.046 kg/s (55,920 lb/h) at a SO_2 feed rate of 0.6101 kg/s (4,842 lb/h), or 11.5 kg steam/kg SO_2 (11.5 lb steam/lb SO_2). Figure 3-16 shows the actual average amount of steam consumed per amount of feed SO_2 for each 30-day period. Actual steam requirements averaged 5% higher than the design expectations. Steam consumption is expected to vary with the absorbing solution feed rate, but the solution feed rate was not measured to confirm this.

During July, the FGD plant operated over a wide range of loads and flue gas flows. Regression analyses were performed on samples of the

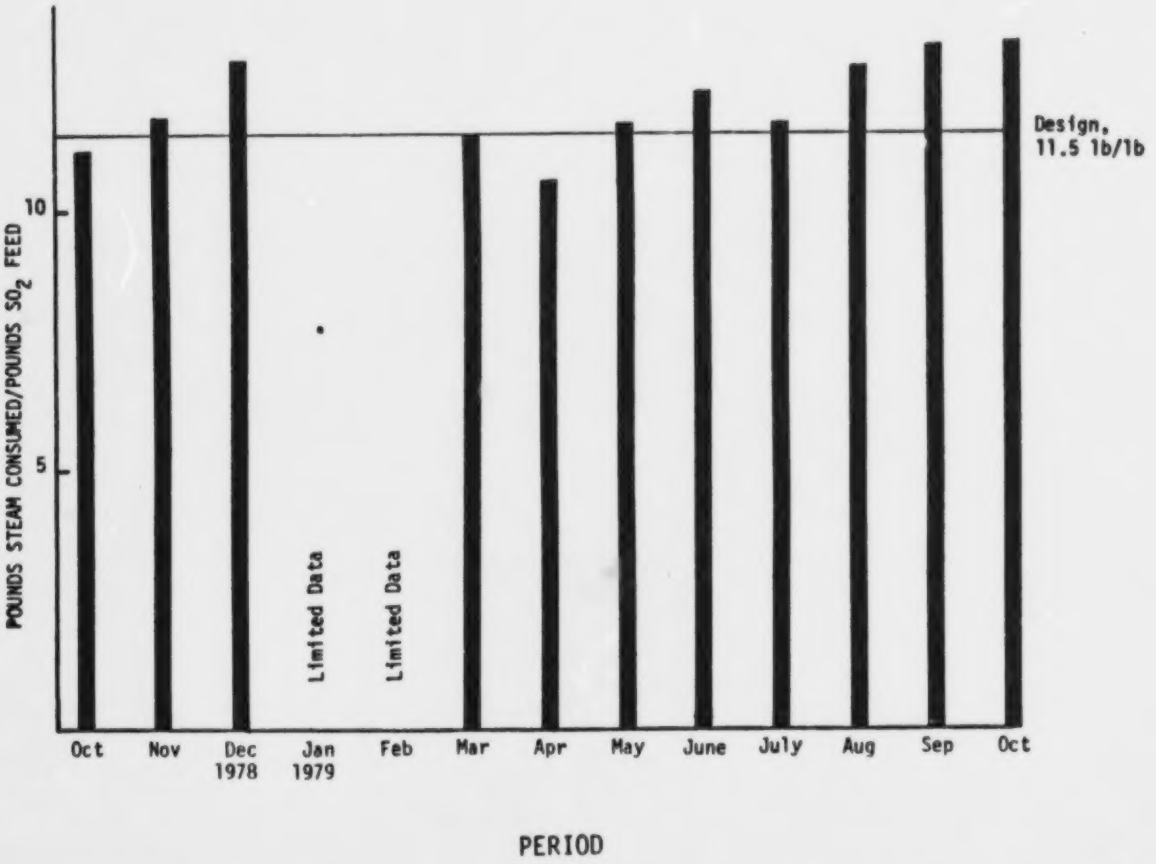


Figure 3-16. Comparison actual steam consumed as a function of SO_2 feed rate with design value.

hourly average data at several load levels. Fairly good linear correlation of steam rate with flue gas flow rate was obtained (Figure 3-17). The correlation coefficient was 0.87 at flue gas flows from 109 m^3/s (231,000 acfm) to 166 m^3/s (352,000 acfm). Flue gas flows typically have exceeded design expectations by about 12%. Attempts to correlate steam with SO_2 feed, with SO_2 removal, and with boiler load were unsuccessful.

The FGD plant consumed a significant proportion of the steam produced by the boiler. Boiler main steam from the superheater section at 12,000 kPa (1800 psi) and 537.7°C (1000°F) was let down and desuperheated to obtain steam for the FGD plant at 3,800 kPa (550 psig) and 398.8°C (750°F). This steam was used in steam turbines to drive the booster blower, the SO_2 compressor and the evaporator circulating pump. However, before the start of the second demonstration year, an electric motor drive for the evaporator circulating pump was installed to eliminate the startup and shutdown problems that occurred when high pressure steam was interrupted by unscheduled boiler shutdowns. The turbine exhaust steam along with additional 3,800 kPa (550 psig) steam that had been let down through a pressure reducing valve was desuperheated further and used for process heat, primarily at the evaporator.

Actual steam consumption at 3,800 kPa, 398.8°C (550 psig, 750°F) averaged 7.03 kg/s (55,800 lb/h) during the second demonstration year and this amount was very close to the expected steam consumption of 7.046 kg/s (55,920 lb/h). In Btu's, this was equivalent to 11% of the boiler input energy derived from fuel, and derated the 115-MW boiler by 8% at the average boiler load of 77 MWe. Table 3-15 summarizes the

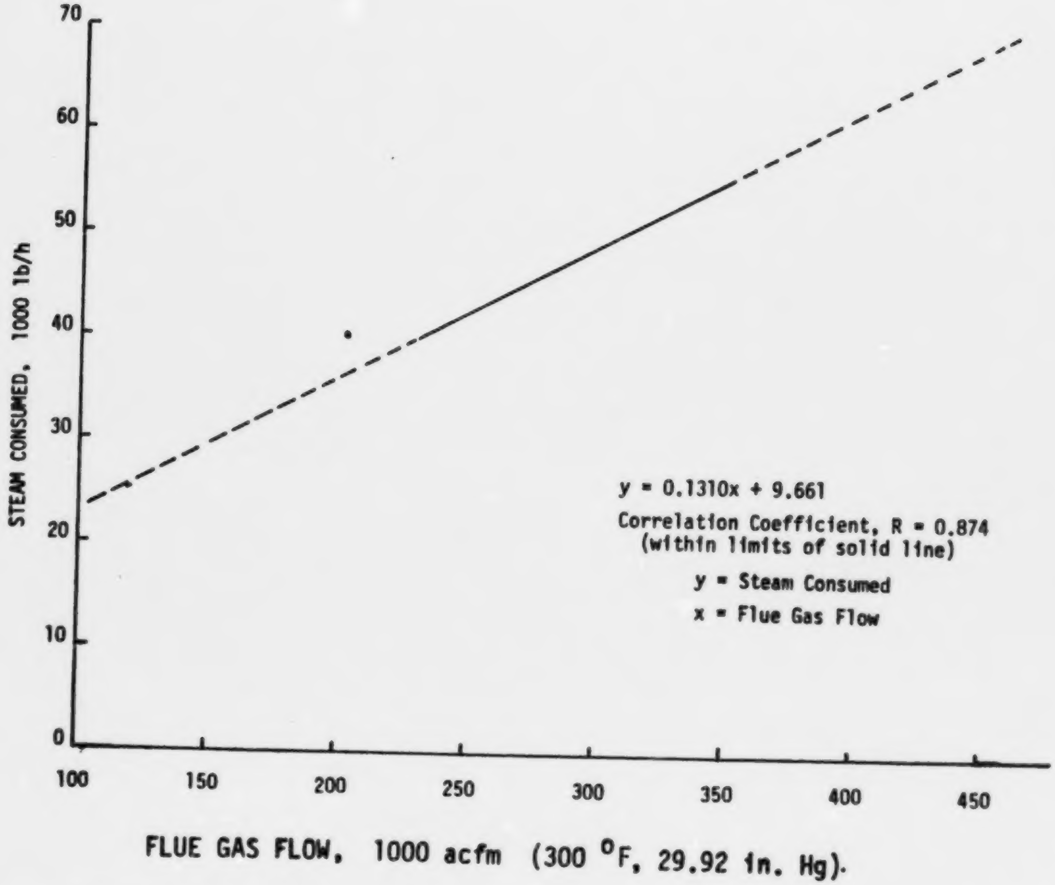


Figure 3-17. FGD steam consumption as a function of flue gas flow.

TABLE 3-15. EFFECT OF FGD STEAM CONSUMPTION ON BOILER

Month	Gross load MWe	FGD steam, kg/s (lb/h)	Energy Equiv. of boiler heat input, J/J (Btu/Btu)	Load Equivalent, MWe	Generating Potential, MW	Boiler Derating, %
Oct. 78	84.7	(57,522) 7.2476	0.101	9.5	94.2	8.2
Nov.	74.0	(51,818) 6.5290	0.105	8.6	82.6	7.1
Dec.	74.7	(57,297) 7.2193	0.117	9.8	83.8	7.2
Jan. 79	78.3	(56,058) 7.0632	0.107	9.4	87.7	7.6
Feb.*						
Mar.	82.6	(54,469) 6.8630	0.097	8.9	91.4	7.9
Apr.	74.0	(51,799) 6.5266	0.105	8.7	82.7	7.2
May	84.7	(59,034) 7.4382	0.104	9.9	94.6	8.3
June	74.0	(50,551) 6.3693	0.102	8.4	82.4	7.1
July	69.7	(51,684) 6.5121	0.112	8.8	78.5	6.8
Aug.	77.2	(56,239) 7.0860	0.109	9.4	86.6	7.5
Sept.	75.1	(65,691) 8.2769	0.133	11.7	86.7	7.5
Oct.	72.9	(60,658) 7.6428	0.126	10.5	83.5	7.2
Average	76.8	(55,800) 7.0307	0.110	9.5	86.1	7.5

*Limited Data.

impact on the boiler of FGD steam consumption for each 30-day operating period.

During Mode 2 operation, steam consumption when the reduction area was not operating or when the bypass damper was open was not much different than during full operation. During complete shutdown, some emergency steam at a lower pressure was used for trace heating and other requirements. This steam, extracted from a turbine stage, was not metered and, therefore, quantities used were not reported.

Electrical Energy

Electric power was metered for all users except the evaporator circulating pump. NIPSCO has estimated the load requirement for the evaporator circulating pump to be 330 kW. For the second demonstration year, metered hourly average electrical energy was 713 kWh, for a total hourly consumption of about 1 MWh. During partial operation of the FGD plant, the electrical energy consumption was virtually identical, and the consumption decreased only slightly with the FGD plant down. Because of the need for continued operation of most electrical equipment (circulating pumps, heaters, evaporator circulating pump, etc.), the hourly average was 673 kWh, which is 94% of the consumption at full operation.

The loss in generating potential from the combined consumption of steam and electrical energy results in a 12% loss of the boiler heat input derived from fuel and electrical energy use by the FGD unit derates the boiler another 1%. Table 3-16 summarizes the impact on the boiler of the combined loss.

TABLE 3-16. EFFECT OF FGD ENERGY CONSUMPTION ON BOILER

Average gross load	76.8 MWe
Average boiler heat input, J/s (Btu/h)	246.479×10^6 (841.586×10^6)
Average energy consumption, J/s Btu/h)	29.408×10^6 (100.410×10^6)
Energy equivalent of boiler heat input, J/J (Btu/Btu)	0.119 (0.119)
Load equivalent	10.5 MWe
Boiler derating	9.1%

Natural Gas

Hourly consumption during Mode 2 operation was $10{,}540 \times 10^6$J (9.994×10^6 Btu), see Table 3-14. Total hourly consumption for the 13 months of the second demonstration year, including all modes of operation and no operation, was $11{,}769 \times 10^6$J (11.163×10^6 Btu) or 12% higher consumption than that for Mode 2 operation (Table 3-17). The additional consumption reflects continuous use during shutdown for operation of the incinerator and for preheating during startups of the reduction unit.

TABLE 3-17. ENERGY CONSUMED AS NATURAL GAS

Total, 13 months, J (Btu)	$59{,}748 \times 10^9$ (56.669×10^9)
FGD Mode 2 operation	5,079 hours
Energy consumed/Mode 2 hour, J/s (Btu/h)	3.2694×10^6 (11.163×10^6)
Boiler operation	8,829 hours
Energy consumed/boiler operating hour, J/s (Btu/h)	1.88×10^6 (6.42×10^6)

PROCESS ECONOMICS

Actual annualized costs, adjusted to projected unit costs and prices, were slightly higher than the projected costs (Table 3-18). Lower utilities and raw material costs that were due to low operability were offset by lower byproduct credits and the other variable costs. The bulk of the other variable costs are for operating labor costs, which continue to accrue while the FGD plant is not operating, and for maintenance. The unit operating cost of 14.9 mills/kWh was 13% higher than the projected unit operating cost. The projected unit operating cost has been adjusted from values used in the interim report[8] to better reflect the FGD design load factor. It is based on the 82.2 MWe of generator output that is possible at the FGD design capacity of 92 MW, and it assumes 351 days of boiler operation per year. Capital cost of the FGD unit and the projected operating costs are shown in Tables 3-19 and 3-20. The unit costs of utilities and raw materials are 1978 costs and, therefore, do not reflect current costs (Table 3-21). Nevertheless, 1978 unit costs have been used again to be consistent with the first demonstration year.

The annual operating costs at a fixed dollar value were about the same for both demonstration years, despite a 54% utilization the second year compared with a 25% utilization for the first year (Table 3-22). Utilization is the hours of FGD operation/hours in a period, expressed as a percentage. The annual operating costs were about the same for both years due to the fixed costs and the operating labor costs that continued to accrue while the FGD plant or the boiler were down. The fixed costs contribute about one-half of the annual operating costs.

TABLE 3-18. ACTUAL ANNUAL OPERATING COST - SECOND DEMONSTRATION YEAR

Item description variable costs	Cost basis		Cost, $	
	Projected	Actual	Projected	Actual
Utilities:				
(a) Steam at \$2.35/1000 lbs	(471,070 x 10^3 lbs) 213,674 x 10^3 kg	(318,608 x 10^3 lbs) 144,518 x 10^3 kg	1,107,000	748,724
(b) Electric power at \$0.016/kWh	7,875,000 kWh	8,079,741 kWh	126,000	129,276
(c) City water		\$7,000 (assumed)	7,000	7,000
Utilities subtotal- -			1,240,000	885,000
Raw Materials:				
(a) Natural gas at \$1.60/10^6 Btu	(130,814 x 10^3 cf, 3704.240 m^3 (1,032 Btu/cf)	(50,508 x 10^3cf, 1,430,227 m^3 (1,025 Btu/cf)	216,000	82,833
(b) Sodium carbonate at \$88.04/ton	(2317 tons) 2,102,000 kg	(2125 tons) 1,978,000 kg	204,000	187,085
Raw materials subtotal- -			420,000	269,918
Sulfur credit at \$35/LT	(7632 LT) 7,754,000 kg	(3252 LT) 3,304,000 kg	(267,120)	(113,820)
Sodium sulfate credit at \$43/ton	(1244 tons) 1,128,000 kg	(308.4 tons) 279,800 kg	(55,980)	(13,878)
By-product credits subtotal- -			(323,000)	(127,698)
All other costs			2,556,000	3,230,503
Total variable costs			3,893,000	4,256,723
TOTAL FIXED COSTS			5,203,896	5,203,896
TOTAL ANNUAL OPERATING COSTS			9,096,896	9,460,619
Unit Operating costs, mills/kWh	6.9 x 10^8 kWh	6.35 x 10^8 kWh	13.2	14.9

TABLE 3-19. CAPITAL COST

Parameters	Cost, $
Direct capital costs	
Absorber & related equipment(1)	7,082,140
Fans(2)	399,130
Reheat(3)	262,550
By-product recovery: purge treatment	1,495,270
By-product recovery: SO_2 reduction	1,143,750
Utilities and services(4)	1,181,040
Stack requirements(5)	146,020
System modifications(6)	241,930
Unidentified	60,000
Direct cost subtotal	11,891,830
Indirect Costs	
Engineering	199,100
In-house construction expense	322,230
Allowance for funds used during construction	775,680
Allowance for start-up	3,700,510
Spares, offsite, taxes, freight, etc.	284,000
Other(7)	958,680
Indirect cost subtotal	6,240,210
Total capital cost	18,132,040
Cost per kilowatt of generating capacity, $/kW	156.85

(1) FGD plant receives flue gas from an existing ESP at a normal dust loading of 0.09 grams/m^3 (0.04 grains/acf). This cost item includes an orifice contactor for additional flyash removal and for cooling and saturation of the gas prior to SO_2 absorption. This cost item also includes all equipment for SO_2 recovery and all equipment for soda ash storage and handling.

(2) Forced draft booster fan.

(3) The natural gas-fired reheater for the absorber exit gas has not been operated due to natural gas restrictions.

(4) Included in this cost item are a 2,000 KVA transformer, natural gas lines, power lines, steam lines, and water lines.

(5) The stack is erected atop the absorber. The top of the stack is 51.2 meters (168 feet) above grade.

(6) Extensive modifications, primarily for winterizing, were made following a winter freeze-up.

(7) Administrative and overhead costs.

TABLE 3-20. PROJECTED ANNUAL OPERATING COST

Parameters Variable Costs	Cost, $
Operating labor	750,000
Maintenance labor and supplies	853,000
Utilities:(1)	
(a) Steam at $2.35/1,000 lb	1,107,000
(b) Electric power at $0.016/kWh	126,000
(c) City water	7,000
(d) Treated water	30,000
Utilities subtotal- - - - - - - - - - -	1,385,000
Raw Materials:	
(a) Natural gas at $1.60/$10^6$ Btu	216,000
(b) Sodium carbonate, 2,317 tons	204,000
(c) Other(2)	86,000
Raw Materials subtotal- - - - - - - - -	506,000
By-product credits(3)	(323,000)
Overhead	837,000
Total variable costs- - - - - - - - - -	3,893,000
Fixed Charges	
Interest	1,925,623
Annual depreciation	1,813,204
Taxes	1,465,069
Total fixed charges - - - - - - - - - -	5,203,896
Total Annual Operating Cost	9,096,896
Unit operating cost, mills/kWh(4)	13.2

(1) No funds included for reheat fuel

(2) Includes operating supplies

(3) Based on 7754 metric tons (7632 LT) of sulfur ($34.45/metric ton ($35/LT) and 1128 metric tons (1244 tons) of sodium sulfate ($49.62/metric ton ($45/ton).

(4) Based on 6.9 x 10^8 kWh. This is based on a projected load factor of 85.9% of a FGD plant design capacity of 92 MW, or a boiler load equivalent of 82.2 MWe.

TABLE 3-21. ESCALATION AND FLUCTUATION OF RAW MATERIAL AND BYPRODUCT COSTS

Material	Projected cost, 1978	April 1979
Soda ash	$88.04/ton	$98.10/ton
Sulfur	$35.00/LT	$34.80/LT
Sodium sulfate	$45.00/ton	$25.92/ton

TABLE 3-22. ACTUAL ANNUAL OPERATING COST - FIRST DEMONSTRATION YEAR

Item description	Cost basis		Cost, $	
	Projected	Actual	Projected	Actual
VARIABLE COSTS				
Utilities:				
Steam at \$2.35/1,000 lbs	$471,070 \times 10^3$ lbs	$224,138 \times 10^3$ lbs	1,107,000	526,724
Electric power at \$0.016/kWh	7,875,000 kWh	2,813,000 kWh	126,000	45,008
City water	\$7,000	\$7,000 (assumed)	7,000	7,000
Utilities subtotal- -			1,240,000	578,732
Raw materials:				
Natural gas at \$1.60/10^6 Btu	$130,814 \times 10^3$ cf, 1,032 Btu/cf	$27,393 \times 10^3$ cf, 1,025 Btu/cf	216,000	44,925
Sodium carbonate at \$88.04/ton	2,317 tons	1,431 tons	204,000	125,925
Raw materials subtotal- - - - - - - - - - - - - - - - - - - -			420,000	170,910
Sulfur credit \$35/LT	7,623 LT	1,456 LT	(267,120)	(50,960)
Sodium sulfate credit at \$45/ton	1,244 tons	282.5 tons	(55,980)	(12,713)
By-product credits subtotal - - - - - - - - - - - - - - - - - -			(323,000)	(63,673)
All other costs			2,556,000	3,440,117
Total variable costs			3,893,000	4,126,086
TOTAL FIXED COSTS			5,203,896	5,203,896
TOTAL ANNUAL OPERATING COSTS			9,096,896	9,329,982
Unit operating cost, mills/kWh	6.9×10^8 kWh	5.9×10^8 kWh	13.2	15.8

With 8% more utilization of boiler capacity during the second year, the unit operating cost was 1 mill/kWh less than the first year. At a boiler nameplate capacity of 115.6 MW, the boiler capacity factor was 58% the first year and 63% the second year.

Raw Material Consumption

Soda ash is used as makeup sodium carbonate for the scrubbing process. Usage is determined by buildup of inactive constituents in the absorber/ evaporator loop, such as sulfate and thiosulfate, that must be purged. Any loss from the system due to leaks also would require soda ash makeup. High soda ash consumption during the first demonstration year were due to leaks at the bottom collector tray of the absorber which were repaired before the start of the second demonstration year.

For the 13 months of the second demonstration year, 2,062,000 kg (2,273 tons) of soda ash were consumed, for an average daily consumption of 8,500 kg (9.4 tons) per day, using the total operating days of the absorber/evaporator as the time base. The performance guarantee for acceptance was 6,000 kg (6.6 tons) per day at the design levels of flue gas and inlet SO_2. The annualized consumption was 1,928 x 10^6 kg (2,125 tons).

Natural gas is used as the reductant for converting the SO_2 to elemental sulfur. It is also the fuel used to incinerate the tail gas emitted from the reduction process. The tail gas is returned to the inlet of the absorber after incineration. It was necessary that the incinerator continue to be operated during shutdowns to destroy the reduced sulfur forms that desorb from the reduction unit refractory materials. Thus, there is a corresponding improvement in unit consumption of natural gas with improvement in FGD plant reliability that results from less downtime of the reduction unit. Table 3-23 shows that slightly over 0.4 m^3 (7 cubic feet) of natural gas was consumed per pound of sulfur produced.

TABLE 3-23. NATURAL GAS CONSUMPTION

Annual consumption, m^3 (million cf)	1,430,000 (50.5)
Heating value, J/m^3 (Btu/cf)	38,160,000 (1,025)
For process use, %	87.9
For incineration use, %	12.1
Average consumption, during mode 2 operation, m^3/s (cf/h)	0.07665 (9,745)
For process use, %	92.5
For incinerator use, %	7.5
Process gas/sulfur produced, m^3/kg (cf/lb)	0.39 (6.3)
Total gas/sulfur produced, m^3/kg (cf/lb)	0.43 (6.9)
Consumption during shutdown, % of total	5.6

Byproduct Production

Sulfur production is given in Table 3-24. Annual production for the first year was 1,480 x 10^6 kg (1,631 tons). Significantly more production in the second year reflects higher utilization of the FGD plant and the reduction unit.

TABLE 3-24. BYPRODUCT SULFUR PRODUCTION

Total production, 13 months, kg (ton)	3,533,000 (3,895)
Annual production, kg (ton)	3,304,000 (3,642)
Daily production rate, actual kg/s (ton/day)	0.1812 (17.26)
Daily production rate, design kg/s (ton/day)	0.2528 (24.08)

Dry sodium sulfate production was only 279,800 kg (308.4 tons) for the 13 months and was way below expectations due to the capacity limitation of the purge dryer.

POLLUTING POTENTIAL

The WL/Allied process is a regenerable FGD process and, as such, is intended to have minimum impact on any of the three receiving media (air, water, ground). The intended and actual impact (other than SO_2 emissions) on the environment are summarized as follows:

- Particulate matter. Inlet and outlet grain loading was about as expected. Particle removal by the FGD plant was 32% during the controlled condition tests. Removal was accomplished by two means as planned: (1) the fly ash prescrubber and (2) the fly ash in-line filter in the absorber/evaporator loop. Purge from the prescrubber and sluicing water from the fly ash filters was sent to the fly ash ponds.

- Nitrogen oxides (NO_x). The fate of the NO_x in the inlet flue gas was not addressed in the initial design information. Actual measurements showed no clear indication of either addition to or reduction in NO_x emissions by the FGD plant. Analysis of the fly ash purge did not show any significant removal by that means.
- Chloride. Substantial quantities of the incoming chloride were removed in the prescrubber. Based on an estimated fly ash purge rate of 442 x 10^{-5} m^3/s (70 gal/min) and average chloride content, the chloride removed is estimated to be about 0.0018 kg/s (14 lb/h). Chloride in the coal averaged 0.0049 kg/s (39 lb/h). Fly ash purge was sent to the fly ash ponds.
- Opacity. Reheat was installed at the absorber outlet to raise the temperature of the exiting gas above saturation and thus make the plume invisible. Due to limited supplies of natural gas, the reheat was not operated and there was a visible plume of water vapor from the stack.
- Sulfate purge. This purge stream was to have been dried and marketed as a byproduct. Not all of the liquid purge could be dried, due to dryer capacity limitations. Thus, some had to be discarded.

BOILER PERFORMANCE FACTORS

Boiler Dependability

The second demonstration year was to have lasted from October 4, 1978 to October 29, 1979; covering thirteen 30-day periods. During the first month (October 1978), FGD plant startup was delayed for seven days to conduct a baseline test (Baseline II). Data from Baseline II is used throughout this report to compare FGD operation with the baseline condition. On October 23, 1979 the boiler was forced down due to turbine blade failures, and remained down for that and for scheduled maintenance. Between October 12, 1978 and October 23, 1979 the boiler was available for FGD plant operation 93% of the time (time that flue gas and utilities were available within design limits). Boiler utilization, total hours boiler operated/total hours in period, was 98%. Boiler utilization for the entire 13-month period was 94%.

The boiler capacity factor was 62.3%, based on the nameplate capacity of 115.6 MW. The capacity factor based on the FGD limitation of 85 MW was 85.3%. Accumulated boiler downtime was a little more than 23 days during the 13-month period. On an annual basis, this is equivalent to 22 days. Projected operating costs are based on 14 days of boiler outage. Figure 3-18 shows the variation in the capacity factor for the 13-month period. Higher factors during January through March reflect low utilization of the FGD plant. For the April through October period of typical FGD plant operation, the boiler capacity factor was 59.8%. Boiler outages are summarized in Table 3-25. The frequency of boiler shutdowns has a significant effect on FGD plant dependability due to the startup time required (see Dependability). The length of boiler shutdowns and their causes are described in Table 3-26.

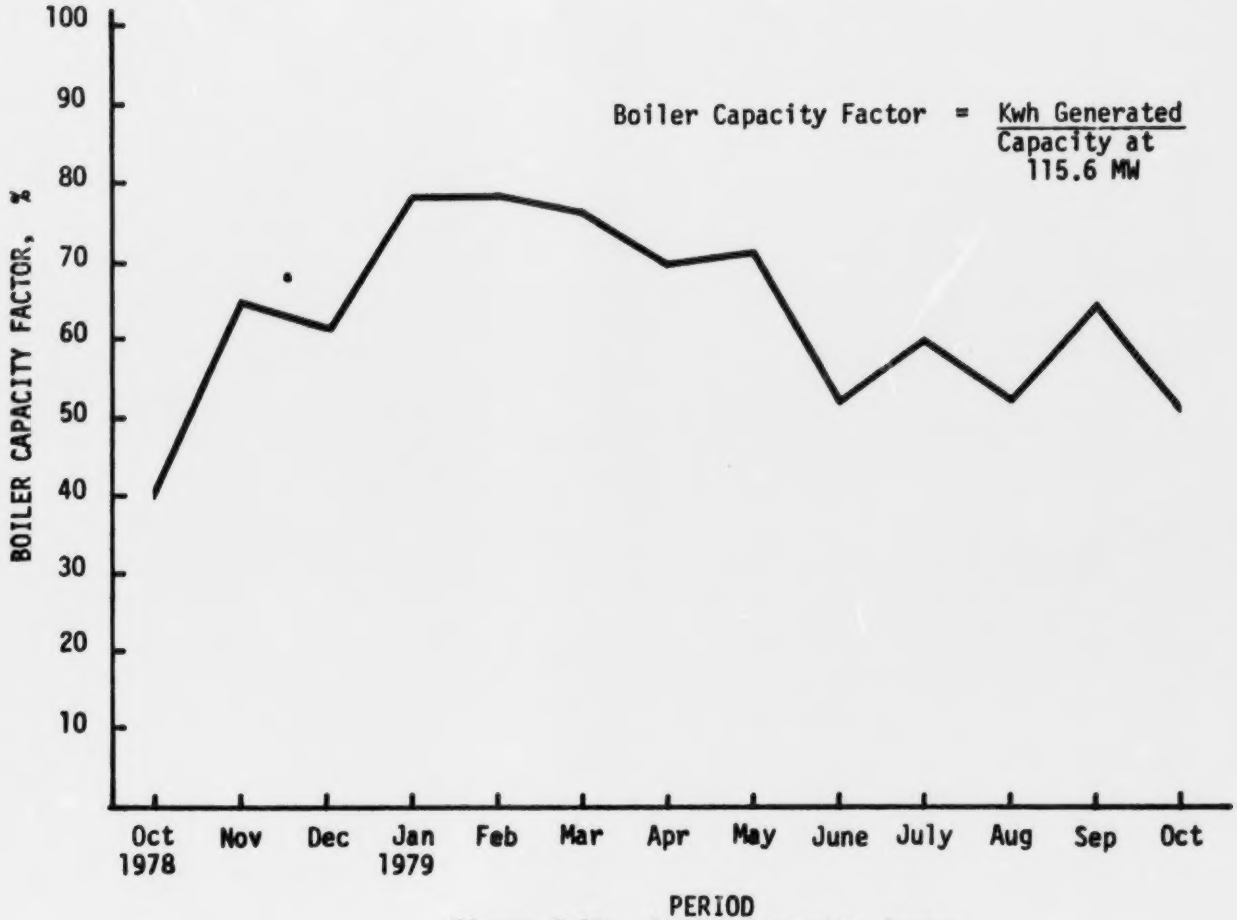


Figure 3-18. Boiler capacity factor.

TABLE 3-25. SUMMARY OF BOILER OUTAGES

Total hours down - 541
Number of outages - 15
Length of outage, range - 1-138 hours

TABLE 3-26. BOILER SHUTDOWNS

Date	Hours down	Reason
10/04	17	Late startup
10/05	21	Valve leak
11/28	3	Modulating servo on intercept valve
12/05	3	High silica in boiler feedwater
12/14	58	Clean air heater
02/24	1	Vacuum pump tripped out
05/12	34	Boiler tube leak
05/31	1	Switch yard problem
05/31	100	Repair turbine stop valve
07/30	3	Switch yard problem
08/14	138	Down for 4KV breaker failure. Water hammer in cold reheat line on startup. Cold reheat line inspected and repaired.
08/26	18	ESP repair
08/30	11	Generator high voltage trip
09/24	3	Switch yard problem
10/23	130	Turbine blade failures

Economics of Boiler Operation

The cost of generating a kWh of electrical energy was higher as a result of the energy taken from the boiler by the FGD plant, primarily as main steam (see Energy Consumption). Figure 3-19 illustrates the increase in heat rates from operation of the FGD plant. In the typical boiler load range of 70 to 85 MWe, heat rates were 10 to 11% higher during FGD plant operation than during Baseline II (Table 3-27). This is in agreement with the equivalent boiler heat input energy of 11% that was calculated from the FGD plant steam consumption. To obtain heat rates during FGD operation, boiler heat input and gross load data during typical operation in April and July were correlated by linear regression analysis (Figure 3-20).

TABLE 3-27. ADDITIONAL HEAT INPUT REQUIREMENTS WITH FGD OPERATION

Load, MWe	Baseline II heat rate 10^3 J/kWh (Btu/kWh)	Heat rate, FGD operating 10^3 J/kWh (Btu/kWh)	Increase in heat input per kWh %
50	10,940 (10,380)	12,290 (11,660)	12
60	10,750 (10,200)	11,960 (11,340)	11
70	10,600 (10,050)	11,710 (11,110)	11
80	10,480 (9,940)	11,520 (10,930)	10
85	10,440 (9,900)	11,450 (10,860)	10

Also affected was the electrical energy available for distribution (net load) as demonstrated by comparing the baseline condition of the Baseline II tests with the controlled condition tests during FGD operation (Figure 3-21).

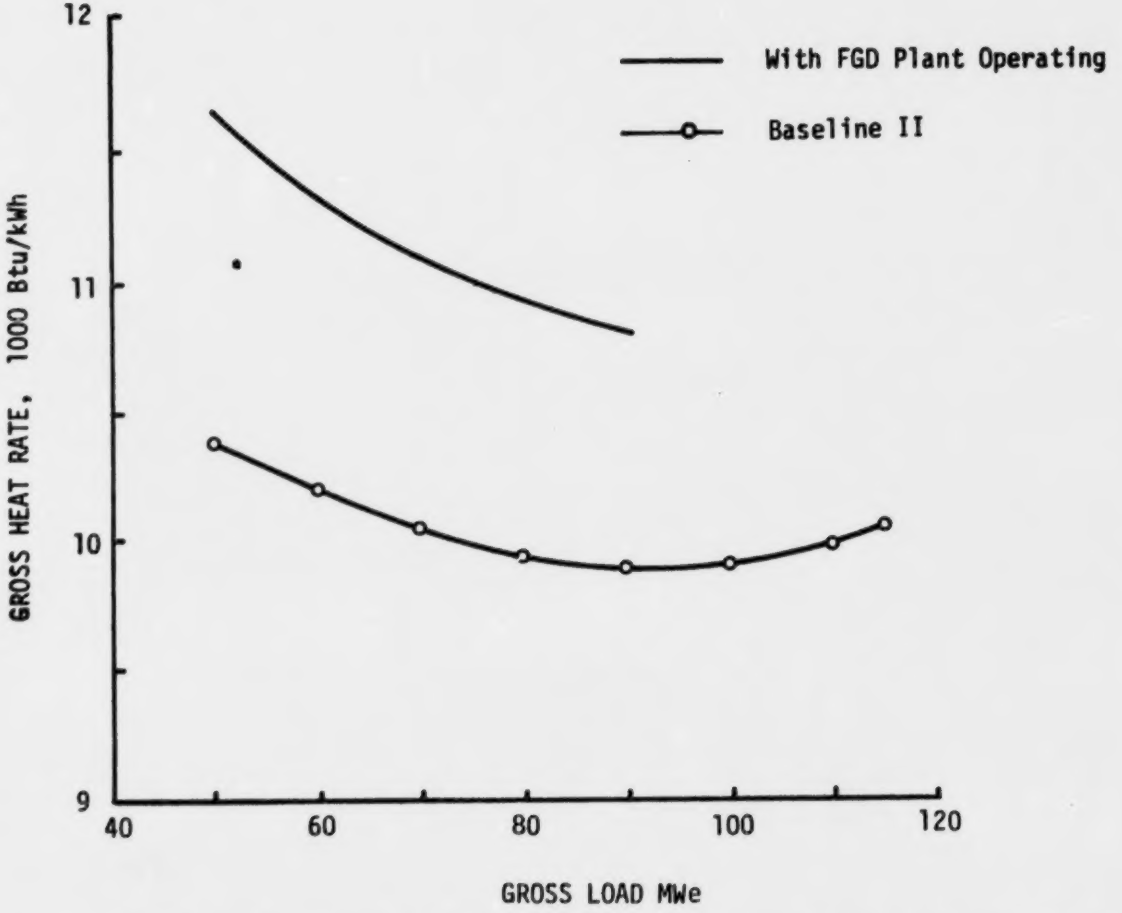


Figure 3-19. Comparison of heat rates with and without FGD operation.

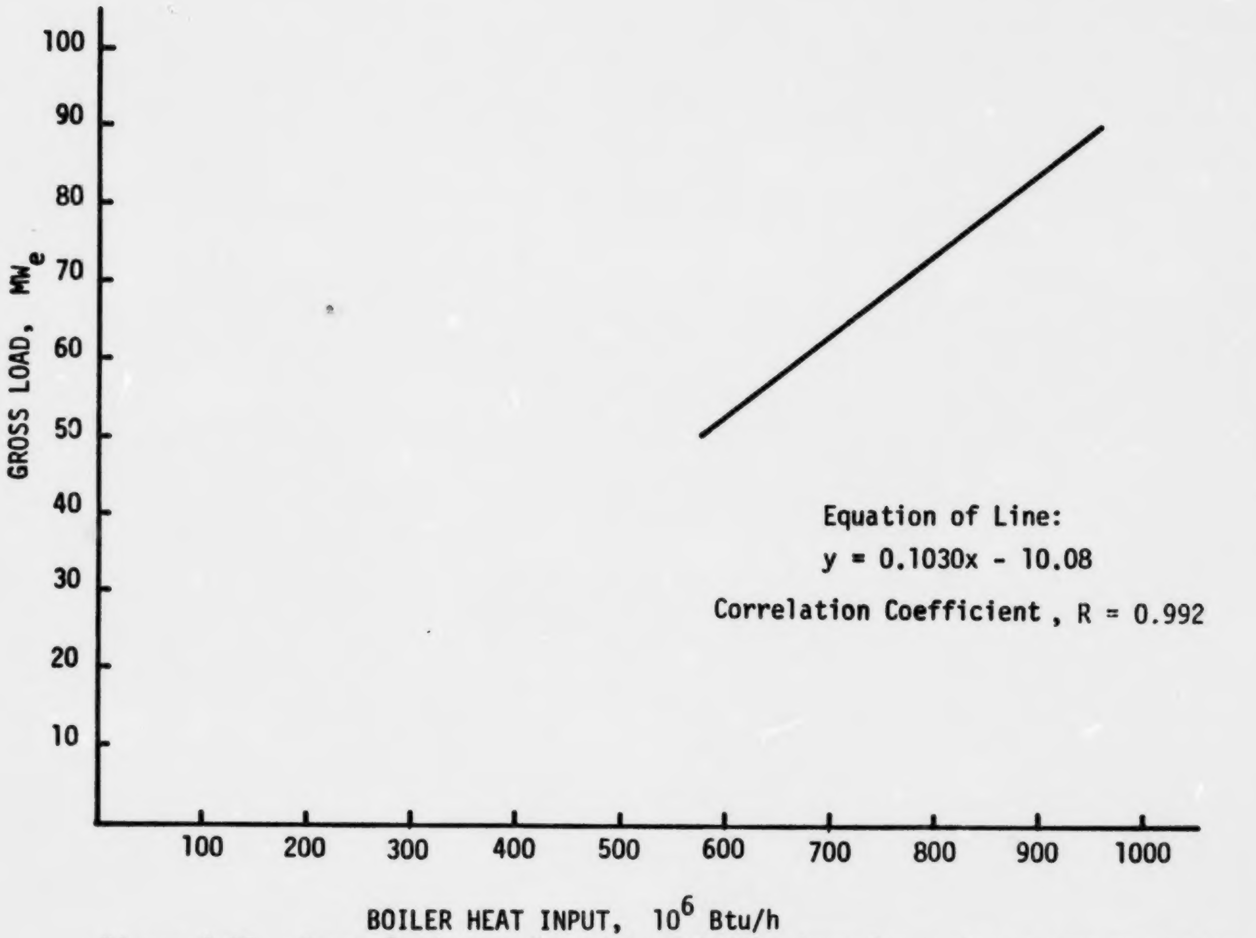


Figure 3-20. Gross load as a function of boiler heat input during FGD operation.

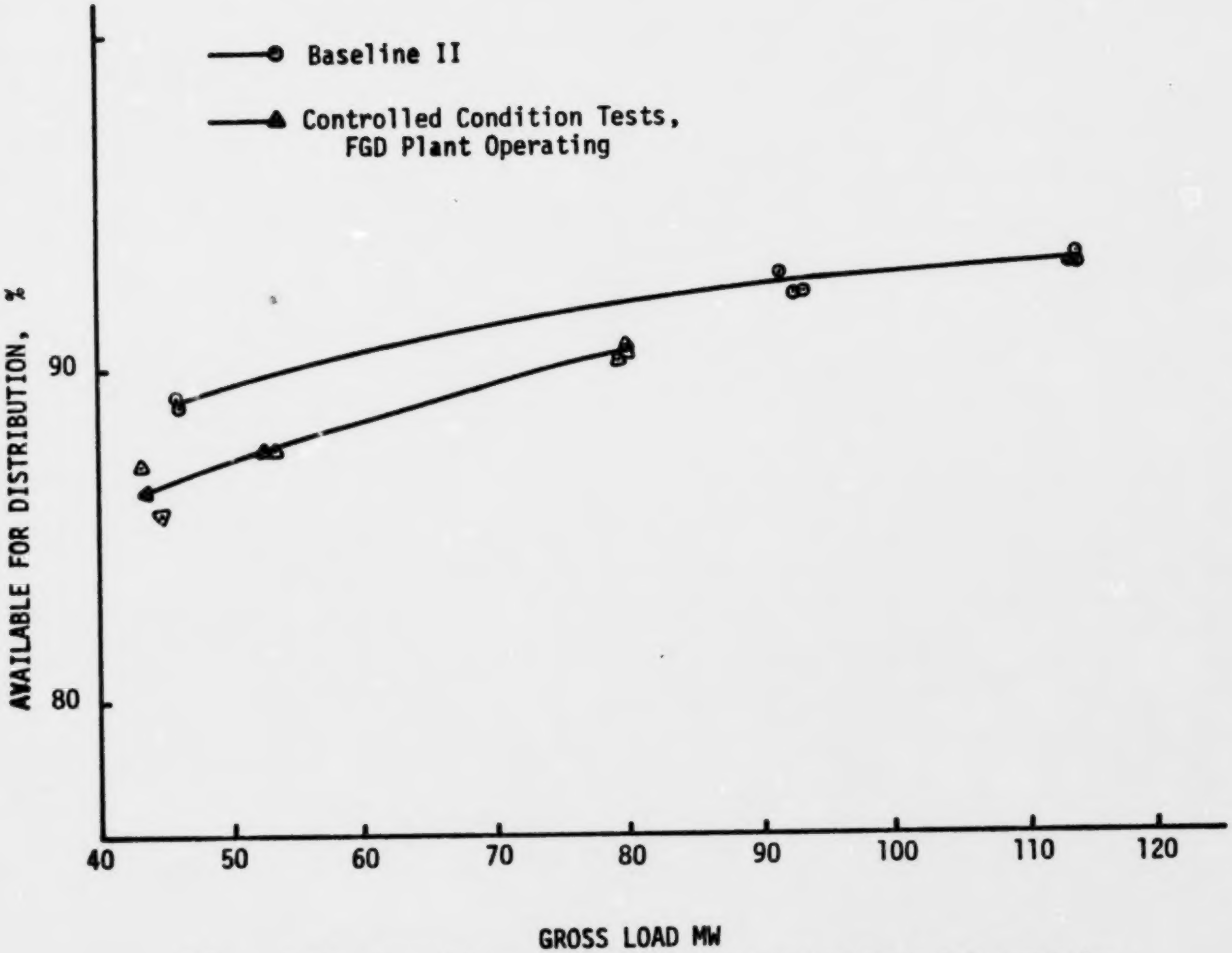


Figure 3-21. Comparison of net load with and without FGD operation.

Flue Gas Quality

Feed characteristics were previously described (SO_2 Removal), particularly with regard to SO_2 feed rate and flue gas flow rate. We will now examine the quality of the flue gas and the fuel in more detail, and report the results of controlled condition tests and how the flue gas characteristics during FGD operation compare with the baseline characteristics.

Fuel Quality--

Coal was supplied from the Captain mine located on the No. 6 bed in Southwest Illinois. The coal is a high volatile bituminous, H.V.C. rank (ASTM D388 designation). The coal was well suited for the Mitchell No. 11 boiler and was within the FGD plant's specification for sulfur content. Table 3-28 shows the average coal composition and variability of the coal as fed to the weigh feeders and before pulverization (raw coal). This summary represents the unweighted average of coal samples collected every sixth day that the FGD plant was operating. Other than some medium sulfur coal burned in July, it is all Captain coal. The sulfur content averaged 3.09 wt.% and had an estimated range of 2.6 to 3.6 wt.% (based on a distribution of two standard deviations from the mean). This was a slightly wider range than the expected FGD plant limitations of about 2.8 to 3.5 wt.%. The ash content averaged 9.8 wt.%. The estimated range was within ±13% of the mean, which indicates a fairly consistent level of ash in the coal. A consistent heating value, estimated range of ±4% of the mean heating value, was partly due to the stable ash content. Moisture content had a somewhat greater variability. The mean value was 13.0 wt.% and the estimated range was ±21% of the

TABLE 3-28. COAL COMPOSITION SUMMARY

Raw, As Received[1]	Average	Std. Dev.	Range $\pm 2\sigma$	Confidence[2] Limits
H_2O, wt%	12.97	1.3474	10.27 - 15.66	12.46 - 13.48
C, wt%	60.93	1.1140	58.70 - 63.16	60.51 - 61.35
H, wt%	4.23	0.1150	4.00 - 4.46	4.19 - 4.27
N, wt%	1.01	0.2036	0.60 - 1.42	0.93 - 1.09
Cl, wt%	0.049	0.0253	0.000- 0.100	0.039- 0.059
S, wt%	3.09	0.2716	2.55 - 3.63	2.99 - 3.19
O, wt%	7.96	0.4780	7.01 - 8.91	7.78 - 8.14
Ash, wt%	9.77	0.6535	8.46 - 11.07	9.52 - 10.02
Heating value, kJ/kg (Btu/lb)	25,482 (10,963)	453 (195)	24,597 - 26,419 (10,582 - 11,366)	25,310 - 25,654 (10,889 - 11,037)

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(1) Coal as fed to the weigh feeders before pulverizing.
(2) 95% confidence limits of the mean.

mean value. The coal composition for Baseline II and for the controlled condition tests are compared in Table 3-28A.

Coal feed rates, coal composition, and flue gas oxygen content were used to calculate flue gas flows. All components of the coal composition except ash were contributors to flue gas volume. The precise flue gas calculation is tedious and had to be done by computer, but a quick estimate was available assuming an average coal composition (Figure 3-22).

Flue Gas Flow Rate--

The normal range of flue gas flows during the second demonstration year was previously defined (see Sulfur Dioxide Removal and Design Limitations).

Excess Air--

Flue gas flow rates were higher than the design level of 151 m^3/s (320,000 acfm) at equivalent generating potentials. The high flue gas flow rates are explained by the higher than expected amount of excess air that can be attributed primarily to inleakage air, believed to be entering at the air preheater. The amount of excess air is controlled to maintain 4% excess oxygen, dry basis, in the flue gas at the inlet of the air preheaters at steaming rates of 80.0 kg/s (600,000 lb/h) and higher, and to maintain increasing concentrations of oxygen with decreasing load below the 80.0 kg/s (600,000 lb/h) steaming rate. For the minimum (46 MWe) operating level, the excess air set point is about 6.2% oxygen in the flue gas. At the 92 MWe through the 115 MWe load levels, the set point is 4.0% oxygen. Baseline tests with the FGD plant down and completely isolated from the boiler were conducted at boiler loads from 46 MWe to 115 MWe. Controlled condition tests during FGD

TABLE 3-28A. COMPARISON OF COAL COMPOSITIONS FOR CONTROLLED CONDITION TESTS WITH BASELINE II TESTS[1]

Composition	Mean		95% confidence limits of the mean	
	Baseline II(2)	CCTP(3)	Baseline II(2)	CCTP(3)
High heating value, kJ/kg (Btu/lb)	25,480 (10,962)	25,429 (10,940)	25,336-25,568 (10,900-11,000)	25,057-25,801 (10,780-11,100)
Moisture, wt.%	12.23	12.68	11.8-12.7	10.2-15.2
Ash, wt.%	10.50	9.86	10.0-11.0	8.9-10.8
Carbon, wt.%	61.19	60.60	60.7-61.6	60.0-61.2
Hydrogen, wt.%	4.16	4.25	4.11-4.21	4.2-4.3
Oxygen, wt.%	8.21	8.29	7.9-8.5	8.0-8.6
Nitrogen, wt.%	0.84	1.05	0.7-1.0	0.9-1.2
Sulfur, wt.%	2.91	3.22	2.9-3.0	3.0-3.5
Chlorine, wt.%	0.04	0.05	0.04-0.04	0.04-0.06

(1) Raw coal.
(2) Baseline tests, October 1978.
(3) Controlled condition test phases, July 1979.

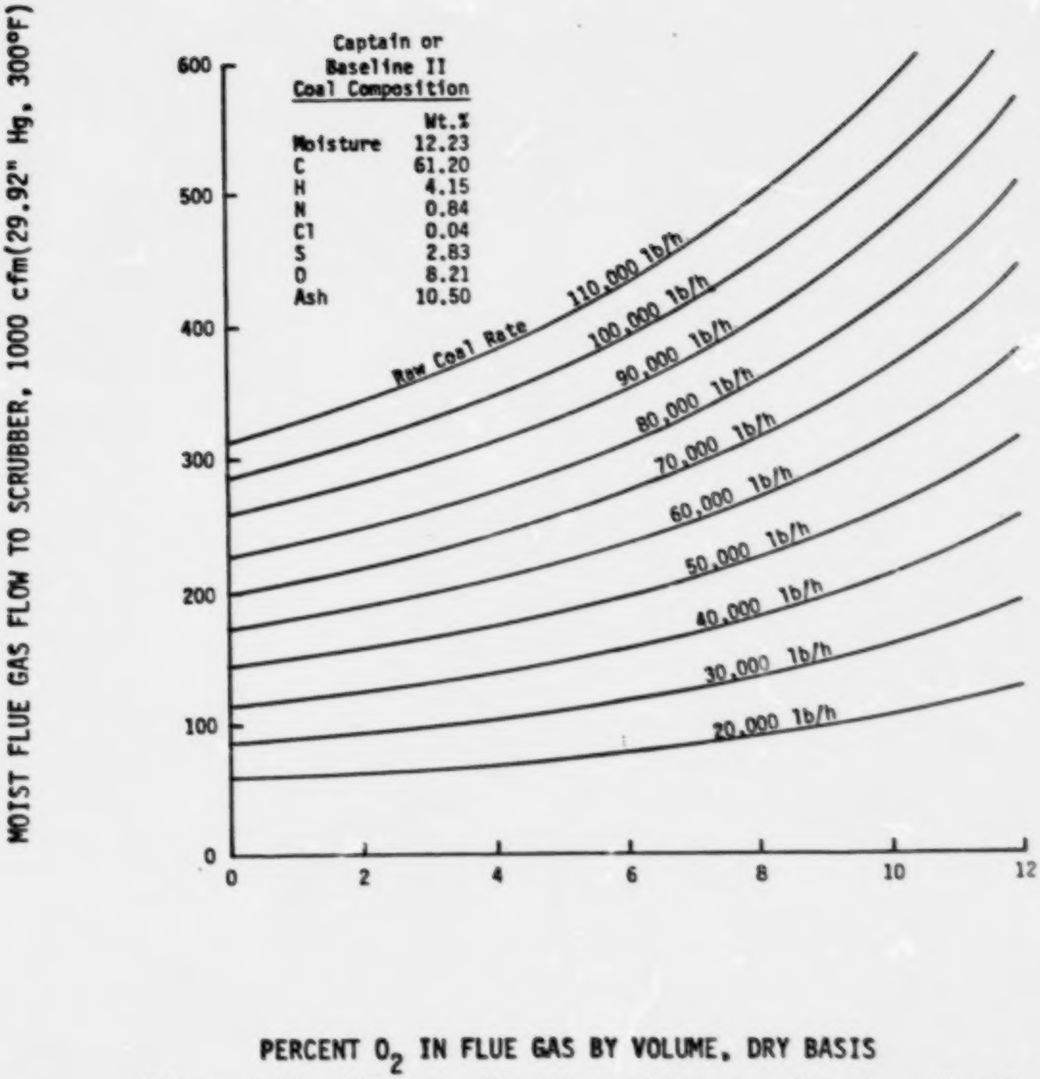


Figure 3-22. Flue gas flow to scrubber as a function of coal rate and scrubber outlet O_2 concentration.

plant operation were conducted at generating potentials from 50 MW to 89 MW (44 MWe to 80 MWe). Table 3-29 compares the baseline and the controlled condition test using excess oxygen data at the inlet and outlet of the air preheaters. Typical variations in the oxygen levels for the second demonstration year are also shown in Table 3-29. Attempts to correlate oxygen with boiler load or coal rates were generally unsuccessful. The typical range of oxygen contents at the inlet of the FGD plant was 5.8 to 8.8 vol.%, although minimum and maximum values overall were 5.6 to 9.9 vol.%. The effect on flue gas flow from the air inleakage across the air preheaters is shown in Figure 3-23.

Pressure and Temperature--

There were high pressure and low pressure limits, respectively, of +10 inches W.C. and -5 inches W.C. for the static pressure in the duct between the ID fans and the booster fan. The bypass damper would open at either limit. The static pressure was used to control the speed of the booster fan. The static pressure during the controlled condition tests was as follows:

• 60 MW generating potential	0.6 kPa (2.5 inches W.C.)
• 88 MW generating potential	0.9 kPa (3.5 inches W.C.)

Inlet temperatures averaged 151.7°C (305°F) during the second demonstration year. However, 38% of the hourly average temperatures were below 148.9°C (300°F) but virtually all of these temperatures were above 137.8°C (280°F). It should be noted that these are single point temperatures well within the flue gas stream, and do not reflect temperatures at and near the duct surfaces. There were, nevertheless, no problems attributable to wet flue gas. The average temperature for each month is shown in Table 3-30.

TABLE 3-29. VARIATION IN EXCESS OXYGEN

Generating Potential, MW	Baseline II O_2, vol.% dry			Generating Potential, MW	CCTP O_3, vol.% dry			Generating Potential, MW	April to August O_2, vol.% dry		
	IN APH,[1]	IN FGD	Δ		IN APH	IN FGD	Δ		IN APH	IN FGD	Δ
46.0	6.4	9.8	3.4	49.9	-	8.3	-	58.1	3.9	8.0	4.1
46.1	7.2	10.5	3.3	50.3	-	10.0	-	65.0	5.1	8.4	3.3
91.7	5.2	7.8	2.6	51.4	-	8.8	-	74.7	5.1	8.8	3.7
92.5	4.4	6.9	2.5	59.8	3.2	7.2	4.0	84.7	3.6	8.8	5.2
93.5	4.3	7.5	3.2	60.7	3.5	7.5	4.0	89.2	3.6	8.1	4.5
113.2	4.0	6.9	2.9	60.8	3.2	9.3	6.1	89.6	3.5	6.4	2.9
113.5	3.8	6.8	3.0	88.3	3.4	7.1	3.7	92.3	4.1	6.8	2.7
114.1	3.6	6.6	3.0	88.6	3.5	6.1	3.6	93.3	3.6	6.7	3.1
				88.8	3.4	6.6	3.2	101.5	5.1	8.4	3.3
								102.3	3.5	5.8	2.3

[1] Air preheater

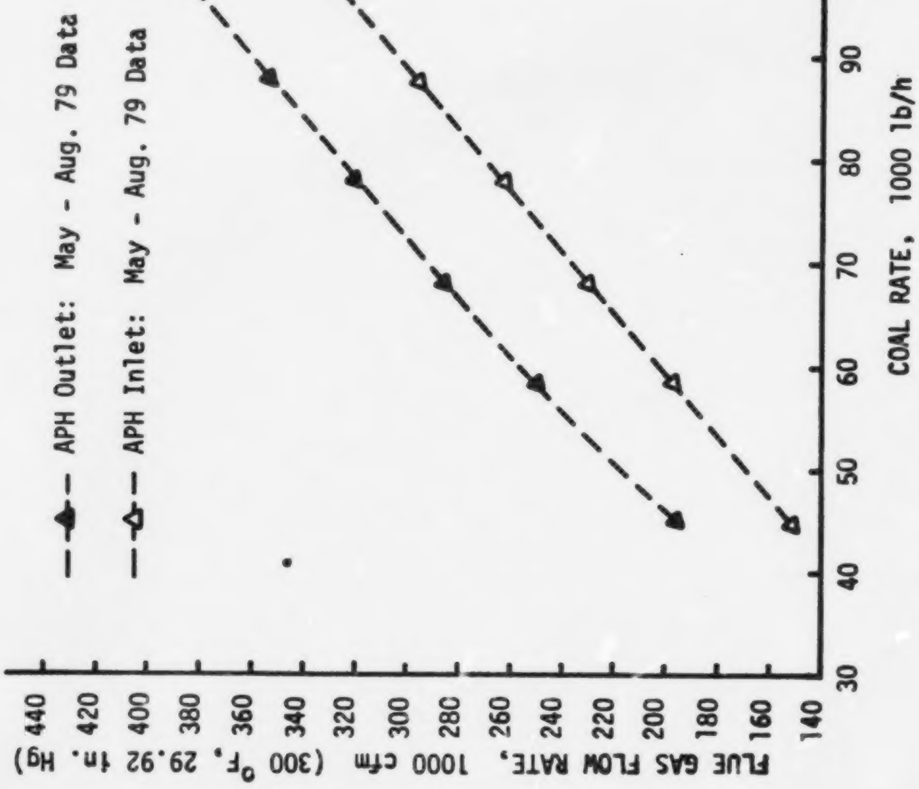


Figure 3-23. Effect of air inleakage on flue gas flow rate.

TABLE 3-30. AVERAGE INLET TEMPERATURE

Month	Temperature, °C(°F)
October 1978	145.5 (294)
November[a]	--
December	147.2 (297)
January 1979	144.4 (292)
February[a]	--
March	157.8 (316)
April	148.3 (299)
May	151.1 (304)
June	156.6 (314)
July	158.9 (318)
August	155.0 (311)
September	152.2 (306)
October	142.2 (288)

[a] Not available.

NO_x Emissions--

The emission rate per boiler heat input is summarized in Table 3-31. Some of the emission rates leaving the FGD absorber indicated NO_x removals of up to 10%. However, these data were offset by some data that showed higher rates at the outlet than at the inlet. Thus, it is unclear that any substantial removal took place.

TABLE 3-31. NO_x EMISSION RATE FROM BOILER

Average	160 ng/J (0.37 $lb/10^6$ Btu)
Standard deviation	±37.4 kg/J (± 0.087 $lb/10^6$ Btu)
Two standard deviations	±74.9 kg/J (± 0.174 $lb/10^6$ Btu)
95% confidence limits of mean	±7.31 kg/J (± 0.017 $lb/10^6$ Btu)

Particle Emissions--

Particle emissions from the ESP measured during the controlled condition tests varied from 30-60 ng/J (0.07 to 0.14 lb/10^6Btu), see Table 3-32. Compared to Baseline II tests, the ESP emissions were lower (Table 3-33).

The ash collected by the ESP had an average unburned carbon content of 6.4 wt.% (determined as loss on ignition) and an average sulfur content of 0.73 wt.%. The unburned carbon was equivalent to about 0.6% by weight of the coal. A value of 0.5 wt.% was assumed to calculate flue gas flow rates.

Ash removal by the FGD plant averaged 32% during the controlled condition tests. Ash removal has been previously described (see Polluting Potential).

TABLE 3-32. PARTICLE EMISSION RATES AND CONCENTRATION

	Test no.					
Parameters	1	2	3	4	5	6
Gross load, MWe	52.4	53.4	53.3	79.6	79.8	79.3
Inlet, kg/m^3 x10^{-3} (gr/acf)	0.057 (0.025)	0.046 (0.020)	0.094 (0.041)	0.055 (0.024)	0.073 (0.032)	0.059 (0.026)
Outlet kg/m^3 x10^{-3} (gr/scf)	0.066 (0.029)	0.050 (0.022)	0.057 (0.025)	0.073 (0.032)	0.078 (0.034)	0.066 (0.029)
Inlet, kg/m^3 x10^{-3} (gr/scf)	0.092 (0.040)	0.073 (0.032)	0.148 (0.065)	0.153 (0.067)	0.112 (0.049)	0.092 (0.040)
Outlet kg/m^3 x10^{-3} (gr/scf)	0.087 (0.038)	0.066 (0.029)	0.076 (0.033)	0.096 (0.042)	0.101 (0.044)	0.085 (0.037)
Inlet, kg/s (lb/h)	0.0085 (67.5)	0.00504 (40.0)	0.0988 (78.4)	0.01564 (124.1)	0.0115 (91.6)	0.00934 (74.1)
Outlet, kg/s (lb/h)	0.583 (46.3)	0.00454 (36.0)	0.00373 (29.6)	0.00908 (72.1)	0.00942 (74.8)	0.00688 (54.6)
Inlet, ng/J (lb/10^6 Btu)	47 (0.11)	30 (0.07)	56 (0.13)	60 (0.14)	43 (0.10)	34 (0.08)
Outlet, ng/J (lb/10^6 Btu)	34.4 (0.08)	25.8 (0.06)	21.5 (0.05)	34.4 (0.08)	34.4 (0.08)	25.8 (0.06)
Ash in raw coal, wt%	9.56	10.13	10.86	9.75	8.79	9.46
Composition of ash collected by ESP						
-Sulfur, wt % dry	1.16	0.63	0.54	0.64	0.72	0.67
-Loss on ignition, wt% dry	7.02	6.67	7.02	5.55	5.81	6.38
Oxygen inlet, vol% dry	7.2	9.3	7.5	6.1	6.6	7.1
Oxygen outlet, vol% dry	7.2	8.2	7.5	6.7	6.6	6.9

TABLE 3-33. PARTICLE EMISSIONS FROM BOILER: BASELINE II AND FGD OPERATION

Boiler heat input, 10^6 J/s (10^6 Btu/h)	Particle emissions, ng/J (lb/10^6 Btu)	
	Baseline II*	CCTP†
140.3 (479.2)	125 (0.29)	-
141.6 (483.6)	95 (0.22)	-
175.4 (599.0)	-	56 (0.13)
179.2 (611.7)	-	30 (0.07)
179.3 (612.1)	-	47 (0.11)
255.0 (870.8)	-	60 (0.14)
259.3 (885.5)	-	43 (0.10)
259.4 (885.8)	-	34 (0.08)
264.5 (903.0)	310 (0.72)	-
265.7 (907.1)	125 (0.29)	0
274.2 (936.4)	95 (0.22)	0

*FGD plant down and completely isolated from boiler, October 1978.

†Controlled condition tests with FGD plant operating, July 1979.

Chloride--

The chloride feed to the boiler and, presumably to the FGD plant, averaged 0.0049 kg/s (39 lb/h) for the second demonstration year (Table 3-34). Analysis of the fly ash purge indicates that substantial amounts of chloride were removed by the prescrubber (see Polluting Potential).

TABLE 3-34. CHLORIDE IN COAL DURING PERIODS OF FGD PLANT FULL OPERATION

Month	Total Cl^-, kg (lbs)	Cl^-, kg/s (lbs/h)
October 1978	1,467 (3,234)	0.0026 (21)
November*	--	--
December	12,333 (27,189)	0.0078 (62)
January 1979	4,466 (9,847)	0.0078 (62)
February*	--	--
March	4,494 (9,908)	0.0067 (53)
April	7,417.1 (16,352)	0.0040 (32)
May	5,939.3 (13,094)	0.0033 (26)
June	5,248.1 (11,570)	0.0047 (37)
July	7,097.8 (15,648)	0.0036 (29)
August	8,864.1 (19,542)	0.0060 (48)
September	8,247.6 (18,183)	0.0045 (36)
October	4,460 (9,834)	0.0030 (24)
Average	--	0.0049 (39)

*Data not available.

Retrofit Effects

The major effect of the retrofit is, of course, the derating of the boiler and the effect on heat rates from FGD plant consumption of boiler input energy. The drain of boiler main steam (plus condensate for desuperheating) by the FGD plant also upsets the turbine steam balance and must affect turbine operation. The effect on turbine performance was not determined. The relative amount of boiler main steam consumed by the FGD plant is shown in Table 3-35.

TABLE 3-35. STEAM DRAIN FROM BOILER BY FGD PLANT

Gross load, MWe	Boiler main steam, 1000 kg/s (1000 lb/h)	% of main steam to FGD plant
54	0.0444 (352)	11.1
56	0.0469 (372)	10.1
61.5	0.0525 (417)	9.9
69	0.0601 (477)	8.7
70	0.0612 (486)	10.1
75	0.0663 (526)	7.7
82	0.0734 (583)	8.3
83	0.0743 (590)	8.4
84	0.0755 (599)	8.7
85	0.0765 (607)	8.7

It is seen that the minimum impact on boiler main steam (and minimum economic penalty) is an 8 to 9% steam drain at the normal operating range of 75 to 85 MWe. The Table 3-35 data represent operation during July 1979. This month was selected for its wide range in steam rates. The gross load was measured directly. A correlation of gross load with boiler main steam was used to determine boiler main steam. Measured FGD steam rates, corrected for condensate addition for desuperheating, were used.

Another possible effect from retrofit would be in boiler draft from a change in back pressure on the ID fans. Figure 3-24 indicates slightly lower boiler resistance during FGD plant operation.

DATA VALIDITY

The credibility of the results that have been reported in the preceding subsections of Section 3 is dependent on the validity of the raw and processed data that was used to obtain the results. Considering the great number of parameters that were being measured at a high frequency of data collection, it was inevitable that some data would be lost or invalidated during test equipment failures. On the other hand, not all of the needed process parameters could be measured, and some indirect means of obtaining this data had to be devised. The most important examples were flue gas flow (calculated from coal composition, coal rate and excess oxygen) and moisture in the flue gas (calculated from vapor pressure data at the outlet, estimated for the inlet). Also, indirect means were used in the event of an instrument failure to optimize data access from the instruments in service. Data validity assurance takes three forms:

1. Quality control for the measurement methods.
2. Minimization of error in results that occur from the accumulation of errors from multiple data sources.
3. Validity of backup systems.

The degree to which each of these contributed to data validity will now be described.

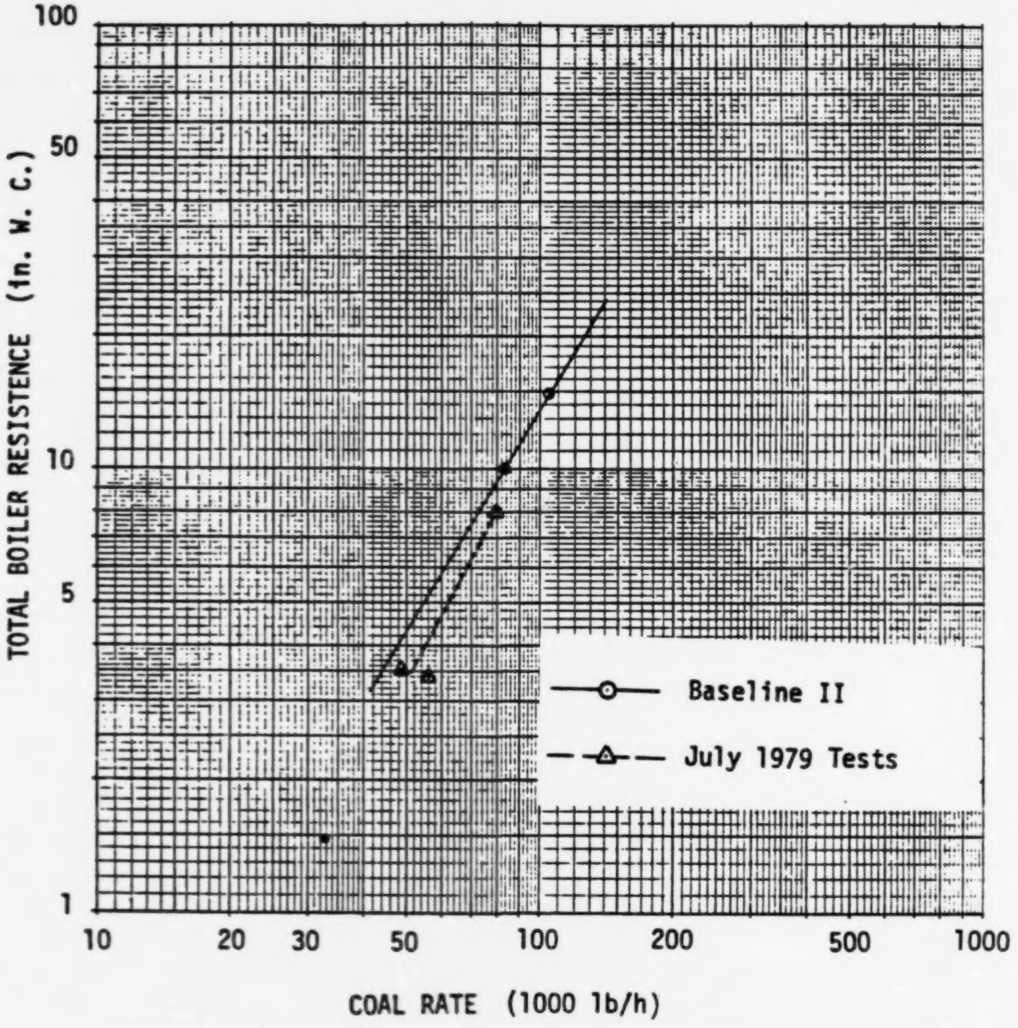


Figure 3-24. Boiler resistance as a function of coal rate.

Quality Control

There were four categories of measurement for which there was a potential need for quality control:

1. Monitoring system serviced by test contractor.
2. Instruments serviced by others (NIPSCO, Allied).
3. Routine manual sampling and analysis.
4. Manual sampling and analysis for special tests and controlled conditions tests.

Monitoring Systems --

Key elements of the quality control program were:

- Daily calibration of the primary instruments (SO_2, oxygen, and carbon dioxide analyzers), including records of drift characteristics.
- Periodic calibration of other monitors.
- Calibration gases traceable to NBS standards.
- Well trained personnel.
- A quick response service agreement for maintenance and repair of the minicomputer and the printer.
- Sample line backflushing each cycle.

From the quality assurance standpoint, the test program suffered from lack of redundancy of critical components, particularly the CO_2/O_2 analyzer sampling system. It also suffered from the inability to service the D/A converter and control unit of the DAS. Also, some data was lost or invalid because the system was not attended around the clock.

Major emphasis was placed on frequent and careful calibration of the gas analyzers. A sample calibration record is shown on the following two

DAILY CAL/DRIFT DATA

DATE__________

CO2 Analyzer

DAS Time Off____________. DAS Time On____________.

DRIFT VALUES

ZERO__________.

SPAN__________.

CAL. VALUES

ZERO__________.

SPAN__________.

O2 Analyzer

DAS Time Off____________. DAS Time On____________.

DRIFT VALUES

ZERO__________.

SPAN__________.

CAL. VALUES

ZERO__________.

SPAN__________.

NO Analyzer

DAS Time Off____________. DAS Time On____________.

DRIFT VALUES

ZERO__________.

GAIN__________.

SPAN__________.

CAL. VALUES

ZERO__________.

GAIN__________.

SPAN__________.

DAILY CAL/DRIFT DATA

DATE___________

S02 Analyzer

DAS Time Off_____________.

DAS Time on____________.

DRIFT VALUES

COARSE ZERO__________.

LOW SPAN_______________.

FINE ZERO____________.

HIGH SPAN______________.

SPAN_________________.

CAL. VALUES

COARSE ZERO__________.

LOW SPAN_______________.

FINE ZERO____________.

HIGH SPAN______________.

SPAN_________________.

H2O ANALYZER

DAS Time Off_____________.

DAS Time On____________.

DRIFT VALUES

ZERO___________.

SPAN___________.

CAL. VALUES

ZERO___________.

SPAN___________.

NOx Analyzer

DAS Time Off_____________.

DAS Time On____________.

DRIFT VALUES

ZERO_________________.

GAIN_________________.

SPAN_________________.

CAL. VALUES

ZERO_________________.

GAIN_________________.

SPAN_________________.

pages. A log was kept of instrument component failures, downtimes and uptimes. This record has been invaluable for identifying valid and invalid data.

Instruments Serviced by Others--

Table 3-36 provides a listing of the instruments for the primary data that were serviced by others.

TABLE 3-36. INSTRUMENTS USED IN TEST PROGRAM, SERVICED BY OTHERS

Instrument	Serviced by
Net and gross kWh	NIPSCO
Coal scales	NIPSCO
Boiler main steam flow	NIPSCO
Oxygen inlet air preheater	NIPSCO
FGD steam flow	NIPSCO
Natural gas flow	Allied

The net and gross load data were taken from the NIPSCO meters through TRW serviced transmitters. The analog signals were digitized and processed on TRW equipment. Difficulties in calibrating the TRW transmitter against the NIPSCO kWh meter required that corrections to the load data be put into the data base. The correction to use was determined from two groups of data:

1. Baseline II tests, October 1978.
2. Controlled condition tests, July 1979.

A regression line and equation were determined on 15 sets of data (Figure 3-25). Thus, valid corrections were made using the following equation:

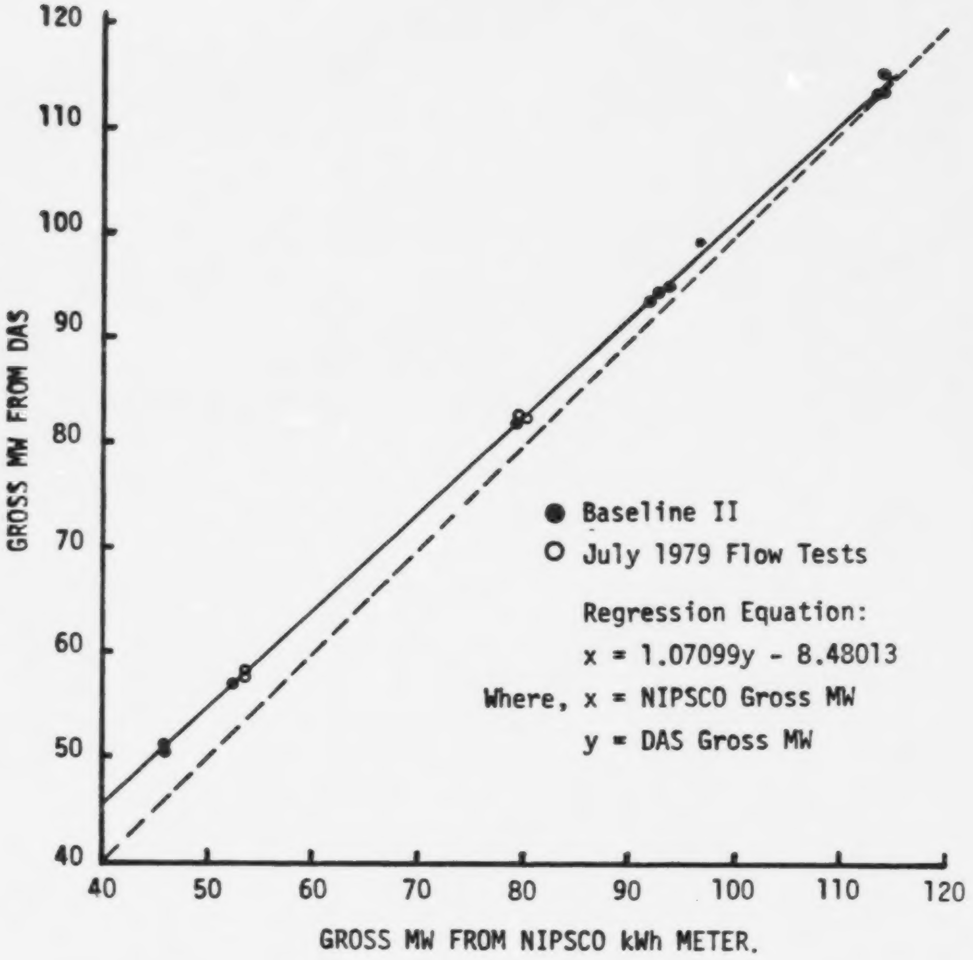


Figure 3-25. Correction factor for DAS gross megawatts.

$$x = 1.07099\ y - 8.48013,$$

where x = NIPSCO gross load in MWe (assumed correct).

y = Test contractor DAS data for gross load in MWe.

The correlation coefficient for the regression line was 0.9999. Since utility company load meters are carefully checked for accuracy, we saw no need for further quality control.

NIPSCO personnel calibrated the coal scales, although infrequently. The scales were weigh feeders that must be kept free of coal dust to remain in calibration. Cleaning of the scales was done routinely. The scales were calibrated with four 25-lb weights. The ASME Power Test Code[11] indicates that the probable coal scale error is ±0.25%. Assuming conservatively as much as ±0.5% error, the uncertainties with three or four coal scales in operation are depicted below:

$$U^2 \text{ (three scales)} = 3 \times 0.5^2$$

$$U = \pm 0.87\%$$

$$U^2 \text{ (four scales)} = 4 \times 0.5^2$$

$$U = \pm 1\%$$

For a typical 10 kg/s (80,000 lb/hr) feed rate, the uncertainty is ±363 kg (800 lb) with four feeders in operation, and ±316 kg (696 lb) with three feeders in operation.

NIPSCO was depended upon for maintaining the accuracy of boiler main steam flow and the oxygen analyzer inlet the air preheaters. The output of the steam flow meter had been corrected for temperature and pressure. The ASME Power Test Code probable error is given at ±1.6% for uncalibrated meters of the differential head type. Estimation of the uncertainty for the oxygen analyzer was not undertaken.

The FGD steam flow signal from the NIPSCO transmitter was accepted by test contractor transmitters and converted to 0 - 1V for input to the DAS. Using NIPSCO pressure and temperature signals, the flow was corrected during data processing. Based on the ASME Power Test Code, the probable error would be ±1.6%.

The natural gas flow was obtained on two flowmeters, for process and for the incinerator. Constant correction factors were used to correct to a STP of 21.1°C (70°F), 101 kPa (14.7 psia).

Routine Sampling and Analysis --

Coal and fly ash purge were sampled routinely. Coal analysis was used to evaluate boiler performance and the backfit effects, as well as to calculate flows and to estimate the SO_2 inlet the FGD plant in case of analyzer failure. Raw coal sampling was done by the NIPSCO coal handlers. They provided the samples in sealed plastic bags that minimize moisture loss. There were no facilities or labor available for obtaining large gross samples and for air drying, as specified in ASTM Method D2234-76,[12] leading to possible large errors in the moisture analysis. Analysis was performed by a commercial laboratory (Commercial Test and Engineering) specializing in fuel analysis. The ASME Power Test Code has provided probable errors for coal analysis.

Manual Sampling and Analysis --

Most of the procedures used were EPA methods and the quality control procedures inherent in these methods were adhered to.

Quality Assurance Audits --

The EPA conducted audits for both continuous monitoring and for manual sampling and analysis. The audit reports had not been released in time to include in this report.

Specific Data Validity

A large part of the data from measurement, and by indirect means, were used for calculating flue gas flows and for determining a sulfur balance across the system. The uncertainty in these results reflects the overall data validity of the test program.

Flue Gas Flow Validation--

To provide the necessary flue gas flow measurements, a calculation procedure was developed. It was validated with actual flow measurements made during acceptance and baseline testing and controlled condition tests. Validation of this method is reported in detail in the Baseline II report.[9] The method uses the ultimate analysis of the coal, coal rate, and CO_2 in the flue gas to calculate a carbon balance as the indirect means for determining the flue gas flow rate:

$$\text{Dry Flue Gas Flow} = \frac{\text{Carbon in coal mole/h}}{\text{CO}_2\text{ in flue gas, vol.\% dry}}$$

The system contains other carbon sources and sinks that affect the carbon balance to a minor extent. These sources and sinks were estimated or ignored, depending upon their size relative to the total carbon in the fuel.

The validity and accuracy of the method has been analyzed in great detail, and the analysis has been reported in the Baseline II report. The estimated uncertainty in flue gas flow has been determined from a sensitivity analysis of the calculation method. Uncertainties in each of the calculation inputs were determined and its effect on flue gas flow was analyzed. The resultant uncertainty in flue gas flow was ±4.2% (Table 3-37).

TABLE 3-37. UNCERTAINTY IN FLUE GAS CALCULATION

Output variable	Uncertainty, %	Effect on flue gas flow, %	
		U	U^2
Coal rate	±2	2.00	4.00
Analysis - H_2O	±8.0	1.14	1.300
C	±1.7	1.31	1.716
H	±2.9	0	--
N	±40	0.38	0.144
S	±17.5	0.37	0.137
O	±11.0	1.11	1.232
Ash	±12.75	1.48	2.190
Carbon in coal not oxidized	±100	0.47	0.221
CO_2 in flue gas	±1	0.95	0.903
SO_2 removed	±3	0.04	0.002
Air humidity estimate	--*	2.36	5.570
Flue gas temperature	±0.7	0.26	0.068
U^2 =			17.483
U =			4.2%

---

*Based on variation in humidity that is due to: air temperature range of 10.0°C - 35.0°C (50°F - 95°F). Relative humidity range of 0% - 70%.

Sulfur Balance Closure --

For the combined period of May, June, July and September sulfur balance closure was as shown in Table 3-38. The method for calculating sulfur balance is given in Appendix E. Greater closure across the boiler plus FGD plant, when determined by flue gas analysis, reflects a sulfur balance closure across the boiler of 103%. This would have to be due to either a positive bias error in the flue gas measurement or a negative bias error in the coal sulfur measurement or some of both. The uncertainty estimate for each of these is as follows.

TABLE 3-38. SULFUR BALANCE DURING MODE 2 OPERATION: MAY, JUNE, JULY, and SEPTEMBER

	Percent Closure	
Description	Across FGD	Across boiler + FGD
Sulfur inlet FGD plant determined from flue gas analysis	82.2	85.2
Sulfur inlet FGD plant determined from coal	84.5	84.8

Table 3-39 gives the error analysis for the coal sulfur measurement. It is seen that the inlet sulfur rate is highly sensitive to small variations in coal sulfur. The uncertainty in coal rate and coal sulfur was taken from the same sensitivity analysis used earlier (see Table 3-37). The uncertainty of sulfur in ash was assumed at ±10%. The uncertainty in inlet sulfur from flue gas analysis is determined as follows:

$$U^2 = 4.181^2 \text{ (flue gas flow)} + 2.5^2 \text{ (ppm } SO_2)$$

$$U = 4.87\%$$

TABLE 3-39. EFFECT OF PROCESS VARIABLES ON SULFUR INLET FGD PLANT - COAL METHOD

		Change in inlet sulfur, %	
Input	Uncertainty,	U	U^2
Coal rate	± 2	± 2	4
Sulfur in coal	±17.5	±17.5	306.25
Sulfur in ash	±10	± 0.36	0.13
U^2 =			310.38
U =			17.6%

Based on this analysis, we have assumed the inlet sulfur determination by flue gas analysis to be more reliable, although the coal method had to be used as backup when the oxygen or CO_2 analyzers were down.

Sulfur balances for each of seven months (April - October) for all operations is shown in Table 3-40. Closure across the FGD plant was calculated by using inlet sulfur determined on the coal, rather than

TABLE 3-40. MONTH BY MONTH SULFUR BALANCE

Month	% Closure, boiler + FGD plant	% Closure, FGD plant
April	87.5	84.0
May	94.6	93.3
June	87.6	90.4
July	87.9	87.3
August	88.7	87.5
September	91.6	89.8
October	87.8	85.2

the flue gas analysis. Table 3-41 compares sulfur balances determined on the coal and flue gas analysis for the month of July, and during full operation only.

TABLE 3-41. COMPARISON OF METHODS FOR DETERMINING SULFUR BALANCE - JULY

	Inlet sulfur determined			
Parameter	Coal		Flue gas analysis	
Sulfur in coal, kg (lbs)	585,587.3	(1,290,999)	585,587.3	(1,290,999)
Sulfur out in ash, kg (lb)	10,924	(24,084)	10,924	(24,084)
Sulfur in inlet flue gas kg (lb)	574,663	(1,266,915)	537,610.9	(1,185,229)
Sulfur out in fly ash purge kg (lb)	11,264	(24,834)	11,264	(24,834)
Sulfur removed, kg (lb)	516,622.2	(1,138,957)	483,312.2	(1,065,521)
Sulfur out stack kg (lb)	46,776.3	(103,124)	43,034	(94,874)
Sulfur to recovery, kg (lb)	458,581.5	(1,010,999)	472,047.7	(1,040,687)
Sulfur out in purge kg (lb)	50,382.3	(111,074)	50,382.3	(111,074)
Sulfur product kg (lb)	350,808	(773,400)	350,808	(773,400)
Total sulfur out kg (lb)	459,231.5	(1,012,432)	455,489.3	(1,004,182)
Closure across FGD plant		79.9%		84.7%

(1) Does not include potential sulfur out in bottom ash and pyrites rejects.

In this report, we have been concerned primarily with the results during full operation. Using the flue gas analysis for inlet sulfur, 15% of the sulfur is unaccounted for during full operation either due to measurement and estimation error or to unknown losses. Sulfur in the fly ash purge is a gross estimate based on a limited amount of purge rate measurements that indicate a sulfur pickup in the prescrubber of 0.00582 kg/s (46.2 lb/h). Sulfur in the purge was not measured because only part of the purge was recovered as dry product. To determine sulfur leaving with the purge, it was assumed that all of the sodium in the makeup soda ash was purged as sulfur forms, with a sodium to sulfur mole ratio of two to one. The purging of significant quantities of sodium thiosulfate ($Na_2S_2O_3$) would introduce an error into this estimate. Based on the error estimates of Table 3-42, the calculated uncertainty is ±7.6% which accounts for only one-half of the unaccounted for sulfur and suggests an unknown loss of about 8%.

TABLE 3-42. ESTIMATED ERRORS FOR CALCULATING SULFUR BALANCE UNCERTAINTY

Parameters	Estimated error, %
Sulfur in flue gas	± 4.87
Sulfur in fly ash purge	±10
Sulfur-in to calculate SO_2-out	±4.87
Sulfur-out to calculate SO_2-out	±4.87
Sulfur in purge	±5
Sulfur product	±5

Given the considerable difficulty indicated in the preceding discussion for defining the exact source or sources of the unaccounted for losses, one can only speculate as to the overall effects on the results of this report. Emphasis on quality control of SO_2 removal performance data rules out any substantial effect on those results.

Backup Methods

The most important backup method employed was a second method for determining the removal efficiency if there was no O_2/CO_2 data and thus no flue gas flow calculations. With SO_2 concentrations being measured, a constant factor was used to correct the outlet concentration to an undiluted basis to determine SO_2 removal. Because the removal efficiency varied only within a narrow range and because there were consistent factors contributing to dilution, possible errors from assuming constant dilutions were not significant. Sulfur balance information also could be obtained without measuring the mass rate of sulfur at the inlet of the FGD plant by using the sulfur to the boiler, corrected for losses to the fly ash removed. It was shown earlier, however, that the uncertainty in the coal sulfur measurement is greater than the inlet sulfur determined from flue gas analysis.

SECTION 4

SPECIAL PROJECTS

As the Wellman-Lord/Allied FGD process demonstration progressed, there were various unanticipated problems and various opportunities for extending our knowledge about how the process works that prompted some special investigations that were not a part of the original test program. These included:

- the effect of upsets and transients on SO_2 removal performance,
- the degree of oxidation during absorption,
- the capacity of the purge treatment dryer, and
- the advantages of increasing the SO_2 removal efficiency of the absorber.

The results of the upsets and transients investigations are included in Section 3 (see Design Limitations). The outcome of the other three studies are summarized in the remainder of this section. Complete results of the purge dryer test and the increased absorption efficiency study are presented in Appendix F and Appendix G, respectively.

SULFATE FORMATION TESTS

The primary objective of these tests was to monitor the amount of sulfate formed in the absorbing solution as a result of oxidation of sulfite and bisulfite. Samples of feed solution and the solution leaving the bottom of the absorber were analyzed to determine the concentrations of the major sulfur forms. From these data and the feed solution flow rate, determination of the rate of sulfate formation was attempted. The determination of specific oxidation rates was complicated by not having flowmeters for all streams and because of the effect of the fly ash prescrubber on the material balance of the system. Figure 4-1 is a simplified flow diagram of the incoming and outgoing streams of the absorber. It was necessary to estimate a water balance to determine the quantity of absorber solution leaving the absorber. Table 4-1 shows the parameters that were measured for each process stream.

TABLE 4-1. SULFATE FORMATION TESTS - SAMPLING AND ANALYSIS SCHEDULE

Sampling Position	Parameter
Flue gas in	Quantity (by pitot traverse or by calculation)
	SO_2 concentration
	SO_3 concentration
	O_2 concentration
	H_2O concentration
Flue gas out	Quantity (by pitot traverse or calculation)
	SO_2 concentration
	SO_3 concentration
	H_2O concentration
Absorbing solution in	Quantity (by flowmeter)
	Na^{+1}
	SO_4^{-2}
	SO_3^{-2}
	$S_2O_5^{-2}$
	$S_2O_3^{-2}$

(continued)

TABLE 4-1. Continued

Absorbing solution out	Quantity (calculated) Na^{+1} SO_4^{-2} SO_3^{-2} $S_2O_5^{-2}$ $S_2O_3^{-2}$
Fly ash purge	Quantity (discharge valve opens on high level in prescrubber sump) SO_4^{-2} SO_3^{-2} $S_2O_5^{-2}$ $S_2O_3^{-2}$
Makeup water	Quantity (by flowmeter)
Boiler	Gross load Coal analysis Coal rate

The sodium forms were analyzed in the FGD plant laboratory and by a subcontract laboratory. A total of seven tests were performed to satisfy the array shown in Table 4-2.

TABLE 4-2. SULFATE FORMATION TEST ARRAY

	Low oxygen	Medium oxygen	High oxygen
High load	X		X
Medium load	X		X
Low load	X	X	

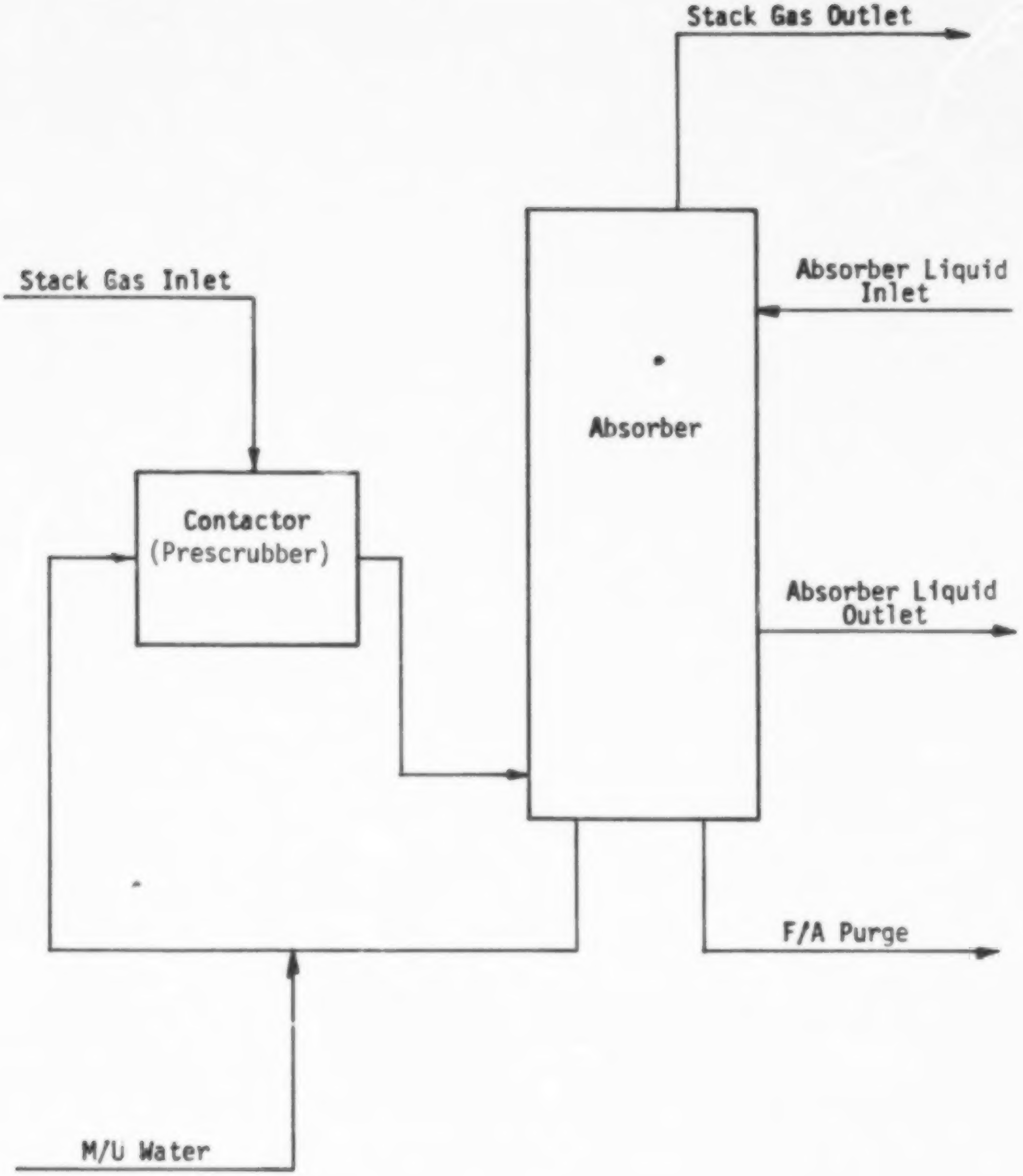

Stack Gas Outlet
Stack Gas Inlet
Absorber Liquid Inlet
Absorber
Contactor (Prescrubber)
Absorber Liquid Outlet
F/A Purge
M/U Water

Figure 4-1. Absorber flow diagram.

Complete results are shown in Table 4-3. The limited testing shows that the sulfate increases significantly during four of the seven tests. The apparent loss of sulfate during the other three tests probably is the result of inaccuracies in estimating flows and in determining the true water balance. Also, a significant increase in the oxidation rate from sulfate formation with increasing excess oxygen (Tests A-1 and A-2) suggests that sulfate formation is a function of the oxygen level in the flue gas. However, additional testing with better metering is needed to confirm this.

PURGE DRYER TEST OBSERVATIONS

The test contractor observed tests performed by Davy McKee. The dryer was tested for approximately one week during May 1979. The tests verified that the dryer did not have heat duty design capacity. The purpose of the test was to demonstrate the heat capacity of the dryer with a water feed onto a sodium sulfate bed in the dryer. If the maximum rate could be reached with water, the dryer was then to be tested with the sodium sulfate solution recovered from the purge solution. Complete results are presented in Appendix F.

BENEFITS OF INCREASING THE ABSORPTION EFFICIENCY

Summary

This section presents the results determined by calculation of a study which defines the maximum practical limits of the SO_2 removal attainable from modifications of the Wellman-Lord FGD process. Specifically, the economic and technical impact and the impact on energy consumption of certain process modifications and equipment changes aimed at maximizing removal efficiency were estimated.

TABLE 4-3. SULFATE FORMATION TEST RESULTS

Test	A-1	A-2	B-1	B-2	I	II	III
Gross MW	87.40	87.12	67.76	69.52	82.69	57.15	80.00
% O_2 (vol.% dry)	7.22	9.46	8.03	9.10	6.89	6.88	6.99
F.G. Inlet							
SO_2 (ppm)	2080	1785	2023	1835	2167.5	2010.5	2065
SO_3 (ppm)	68.3	39.3	65.8	44.3	-	0.0	7.3
F.G. Outlet							
SO_2 (ppm)	213	232	139	138	215	218.7	228.2
SO_3 (ppm)	.06	.09	Lost sample	.08	-	2.3	1.2
Absorber inlet, m^3/s (gpm)	896×10^{-5} (142)	946×10^{-5} (150)	826×10^{-5} (131)	814×10^{-5} (129)	871×10^{-5} (138)	544×10^{-5} (86.3)	845×10^{-5} (134)
S.G.	1.302	1.284	1.291	1.290	1.292	1.292	1.292
% Na_2SO_4	10.13	9.36	10.00	10.33	6.92	6.79	6.08
% $Na_2S_2O_3$	<.01	<.01	<.01	<.01	0.39	0.34	0.39
% $Na_2S_2O_5$	5.01	5.01	4.62	4.75	6.05	5.50	5.32
% Na_2SO_3	16.37	16.22	16.06	16.06	14.07	15.82	15.92
% Na	9.52	9.74	8.60	9.61	-	-	-
% S	8.13	7.92	7.89	8.12	-	-	-
Absorber outlet, m^3/s (gpm)	910.4×10^{-5} (144.3)	892.1×10^{-5} (141.4)	827.7×10^{-5} (131.2)	854.9×10^{-5} (135.5)	891.5×10^{-5} (141.3)	628.0×10^{-5} (99.5)	882.0×10^{-5} (139.8)
S.G.	1.302	1.307	1.300	1.310	1.305	1.305	1.305
% Na_2SO_4	10.46	11.85	9.58	8.38	4.82	6.76	7.20
% $Na_2S_2O_3$	<.01	<.01	<.01	<.01	<0.32	<0.39	<0.31
% $Na_2S_2O_5$	16.36	16.23	15.70	15.70	22.32	18.97	17.60
% Na_2SO_3	8.50	8.34	9.92	9.92	3.19	5.07	6.83
% Na	9.37	9.03	9.08	9.01	-	-	-
% S	10.09	10.26	9.77	9.70	-	-	-

(continued)

TABLE 4-3 (continued)

Test	A-1	A-2	B-1	B-2	I	II	III
Fly ash purge, m^3/s (gpm)	565×10^{-5} (89.5)	357×10^{-5} (56.6)	672.5×10^{-5} (106.6)	561×10^{-5} (88.9)	264×10^{-5} (41.9)	424×10^{-5} (67.3)	290×10^{-5} (45.9)
S.G.	1.000	1.003	.998	1.001	1.0	1.0	1.0
% Na_2SO_4	0.132	0.266	0.135	0.132	0.17	0.00	0.00
% $Na_2S_2O_3$	0.000	0.000	0.000	0.000	0.00	0.00	0.00
% $Na_2S_2O_5$	0.000	0.000	0.000	0.000	0.11	0.12	0.29
% Na_2SO_3	0.000	0.000	0.000	0.000	0.00	0.00	0.00
% Na	0.0105	0.0206	.327	0.296	-	-	-
Make up water, m^3/s (gpm)	890×10^{-5} (141)	890×10^{-5} (141)	890×10^{-5} (141)	890×10^{-5} (141)	770×10^{-5} (122)	770×10^{-5} (122)	852×10^{-5} (135)
% Na	.000726	-	-	-	-	-	-
Change in Sulfur							
#S Absorbed/min.	31.77	30.93	30.56	30.87	37.91	31.21	[illegible].91
$\Delta S\text{-}Na_2SO_4$ (#/min)	1.73	7.28	-1.07	-5.92	-6.48	2.26	4.91
$\Delta S\text{-}Na_2S_2O_3$ (#/min)	0	0	0	0	-0.34	0.42	-0.37
$\Delta S\text{-}Na_2S_2O_5$ (#/min)	60.34	51.18	53.91	63.54	85.38	51.97	64.36
$\Delta S\text{-}Na_2SO_3$ (#/min)	-30.29	-33.52	-21.58	-32.23	-40.69	-23.42	-32.00
% Oxidation $\frac{\Delta S\text{-}SO_4}{\Delta S\text{-}721.14}$	5.45	23.53	-3.50	-17.14	-17.09	7.23	13.30
Inlet F.G. Flow, 148.9^{0}C (300^{0}F), m^3/s	8318.51	9665.19	7767.96	8776.61	9528.62	8557.35	9865.38
(ft^3/min)	287,409	341,323	274,323	309,943	336,500	302,200	348,392
Outlet F.G. flow 54.4^{0}C, (130^{0}F), m^3/s	6488.13	7839.66	6145.78	6957.22	7693.23	6829.00	7992.34
(ft^3/min)	229,126	276,855	217,036	245,692	271,684	241,164	282,247
% Removal	90.09	87.28	93.34	92.65	90.08	89.01	88.90
% Oxidation (SO_3)	3.53	2.46	3.37	2.54	-	-.128	.005
ΔH_2O, m^3/s (gpm)	325×10^{-5} (51.5)	532×10^{-5} (84.4)	217×10^{-5} (34.4)	329×10^{-5} (52.1)	505×10^{-5} (80.1)	346×10^{-5} (54.8)	562×10^{-5} (89.1)

51 ng/J (0.12 lb/10^6 Btu) to 140 ng/J (0.33 lb/10^6 Btu) for the same range of coal sulfur content. For 3.15% sulfur in the coal, the SO_2 emission rate was 110 ng/J (0.25 lb/10^6 Btu) versus 260 ng/J (0.60 lb/10^6 Btu) for 90% SO_2 removal efficiency.

To achieve the benefits of the higher efficiency absorption, more energy must be consumed. For a coal sulfur level of 3.15% the energy requirements increased 17% chiefly as a result of increased heat input for SO_2 recovery and reduction of the SO_2 to sulfur. Annualized costs for four-stage operation were 4% higher than for three-stage operation.

Effect of Partial Flue Gas Treatment--

The second objective of this investigation was to determine the advantages to be gained from high-efficiency absorption when utilized in conjunction with partial treatment of the flue gas. With this operating mode, energy consumptions and costs will be reduced or moderated through reduced throughputs and reduced reheat energy requirements. To reduce the reheat requirement, advantage is taken of a temperature of the bypassed flue gas about 76.6°C (170°F) higher than the absorber outlet temperature. The situation where only enough flue gas was treated to meet the New Source Performance Standard (NSPS) of 520 ng/J (1.2 lb SO_2/10^6 Btu) was considered.

The removal efficiencies required to meet the NSPS are less than 90% for all coal sulfur levels up to and above 4.5%. Therefore, compared to a 90% removal efficiency base case, high-efficiency absorption will only provide part of the benefits if the application of partial treatment is applied just to meet the NSPS. It is apparent that the inlet SO_2 levels will have a profound effect on the amount of flue gas that must be treated. At a 90% SO_2 removal efficiency, the amount of flue gas that must be treated varies asymptotically from 36% for a 1% coal sulfur level to 95% for the 4.5% coal sulfur level. The amount of flue gas which must be treated for a 3.15% sulfur level is 88% and is reduced to 83% with four-stage absorption.

High-efficiency absorption with a fourth stage will also provide an amount of reheat for a recombined gas stream that is comparable to the amount of flue gas that can be bypassed. For an absorber outlet temperature of 52.8°C (127°F), the combined gas stream can be increased to 69.4°C (157°F) for the 3.15% coal sulfur level case. If it is assumed that the exhaust temperature must be 82.2°C (180°F) to provide adequate plume dispersion and to eliminate the visible plume, the amount of reheat energy required is only 43% of the reheat energy required for full flue gas treatment. No reheat is required up to a coal sulfur level of about 2%.

Energy consumed in the FGD plant with partial flue gas treatment is 76% of the energy consumed with four-stage full-treatment operation. It is 88% of the energy consumed with three-stage operation at a 90% SO_2 removal efficiency (with 3.15% sulfur in coal).

The reductions in overall costs attributable to partial treatment are not substantial for coal sulfur levels from 3.15% to 4.5%. At the 3.15% coal sulfur level, annualized costs are 92% of costs for four-stage full-treatment operation and 96% of costs for three-stage operation. However, at the 1% coal sulfur level with substantially less of the flue gas to be treated, the reduction in annualized cost is 28% when going from three stages and 90% removal to four stages and partial treatment. Cost estimates were repeated for a boiler size of 500 MW and are discussed in this report.

Effect of Solution Rerouting--

As an approximation, absorption efficiencies will increase from 95.6% for conventional four-stage operation to about 97% for operation with the sodium carbonate makeup fed directly to the fourth (top) stage and with the rest of the feed solution going to the third stage. A complete discussion of the study is given in Appendix G.

SECTION 5

EVALUATION METHODS

EVALUATION GOALS

Evaluation in response to the test objectives proceeded in six steps:

1. Collect applicable data and operating information.
2. Define hours of operation within each operating mode.
3. Process the raw data and accumulate for each 30-day elapsed period and for specific periods according to the mode of operation.
4. Assess performance with regard to pollutant removal, dependability, energy consumption, and costs.
5. Assess the response of selected dependent variables to changes or fluctuations in the major independent variables.
6. Assess the effect of upsets and transients on SO_2 removal capability.

The evaluation goals were dependent on a variety of measurement techniques that provided the basis for reporting SO_2 removal efficiency, operating load, FGD energy consumption, and cost of utilities. In addition, manual records were used to establish bulk materials consumption and byproduct production. The operating status of the boiler and the FGD plant was an equally important evaluation goal, leading to some rather detailed determinations of the dependability of the two units.

THE TEST SYSTEM

The core test system (Figure 5-1) consisted of sensors for various boiler and flue gas operating variables (with emphasis on the FGD inlet and outlet flue gas parameters) and accumulation of the sensor analog signals by a data acquisition system (DAS). The frequency of analog signal scan by the DAS was three or six minutes, from which one-hour averages were computed. The DAS had the capability for storing the data on magnetic tape; however, hardware difficulties with the tape transport unit were experienced such that very little automated data reduction was possible. Backup storage was available on teletype printouts or on charts taken from strip chart recorders. These data sources had to be utilized at considerable penalty in the time required to access the data and reduce it manually.

It was essential that the test system data be correlated with operational disruptions or limitations. Daily meetings were held with NIPSCO and Allied Chemical representatives to receive reports on the operating status of the boiler and the FGD plant. Use was also made of NIPSCO and Allied reports to obtain raw material rates, product rates and costs.

Table 5-1 is a matrix showing the parameters to be measured at each sampling position. The numbered data items represent the DAS data channels sampled every three or six minutes. The X's indicate less frequent sampling at frequencies of every 24 hours, every 6 days, every 30 days, or for special tests, at least once during the test program.

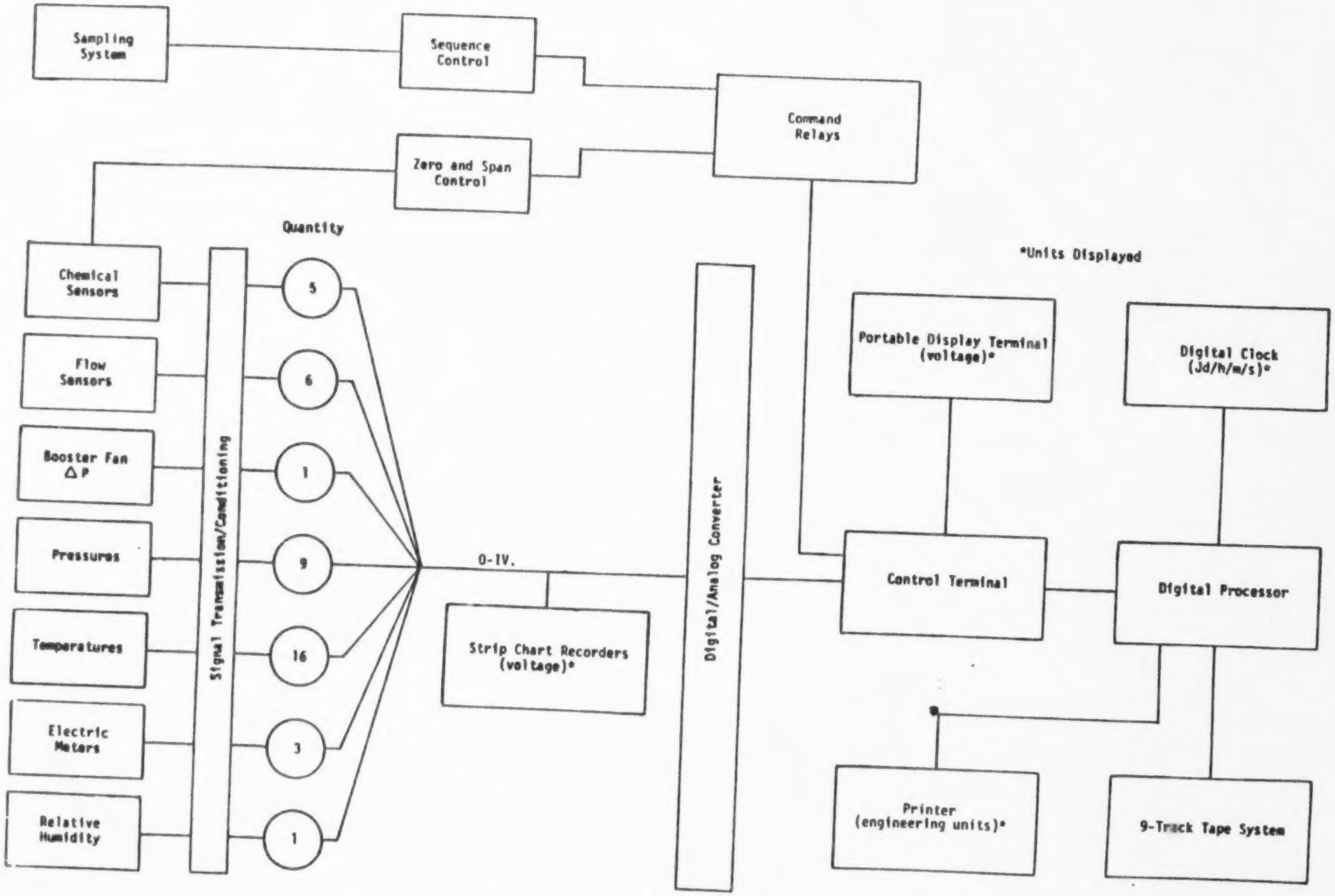

Sampling System
Sequence Control
Command Relays
Zero and Span Control
Quantity
*Units Displayed
Chemical Sensors
Flow Sensors
Booster Fan ΔP
Pressures
Temperatures
Electric Meters
Relative Humidity
Signal Transmission/Conditioning
5
6
1
9
16
3
1
0-1V.
Strip Chart Recorders (voltage)*
Digital/Analog Converter
Portable Display Terminal (voltage)*
Digital Clock (Jd/h/m/s)*
Control Terminal
Digital Processor
Printer (engineering units)*
9-Track Tape System

Figure 5-1. Block diagram of test system.

TABLE 5-1. TEST PARAMETERS

Sampling Position Parameters	Raw Coal 1	Coal As Fired 2	Ambient Air 3	Combustion Air to Boiler 4	Fly Ash (Special) 7	Steam Drum 8	Secondary S.H. Outlet 9	NIPSCO Load Gross 13A	NIPSCO Load Net 13B	Flue Gas Inlet APH 14	Booster Fan Inlet 15	Booster Fan Outlet 15A	Booster Fan 15B	Absorber Outlet 16	Fly Ash Purge 17	Soda Ash Feed 19	Total Natural Gas 21	Purge Solids 22	Sulfur 23	Demo Steam 24	Demo Electricity 25	Boiler Gas Side 33	Auxiliary Amperages 34
1. SO_2											00			00									
2. SO_3											X			X									
3. Particle Mass												X		X	X								
4. Particle Size												X		X									
5. Proximate	X	X																					
6. Ultimate	X	X																					
7. H_2O																							
8. Trace Metals																							
9. Chloride	X	X			X										X								
10. Pressure						07	08										X			X		X	
11. Volume																	X			X			
12. Temperature		X	X	X			49			X							X			X			
13. Weight	X	X			E										E	X		X	X				
14. CO_2											02			02									
15. O_2										03	04			04									
16. Booster Fan ΔP											17	17											
17. NO_x											06			06									
18. Booster Fan RPM													18										
19. Total Sulfur	X	X													X								
20. Heating Value	X	X															X						
21. Combustibles/Sulfur					X																		
22. Relative Humidity				X																			
23. Barometer				X																			
24. Sulfur Species																							
25. Total N															X								
26. COD															X								
27. Gross kWh								38													X		
28. Net kWh									39														
29. Voltage																							
30. Current																							X
31. Soot Blower Status																						X	

Notes: (1) Numbered items of data denote DAS channels.
(2) X's denote non-DAS sampling.

The test system in the following description is the system in place at the completion of testing. Minor improvements and changes were made throughout the test program. The effects, if any, of these changes on the test results have been described in Section 3 under Data Validity.

Sensors and Analog Signal Transmission

The test system was designed for one signal transmission level of 0-1 volt. Thermocouples for temperature measurement and some of the other sensors required signal conditioning to conform. Table 5-2 is a list of the channels for which data was transmitted to the DAS.

TABLE 5-2. TEST SYSTEM CHANNELS

Channel No.	Parameter	Reported as
00-I	SO_2 inlet	ppmv
00-0	SO_2 outlet	ppmv
02-I	CO_2 inlet	vol. %
02-0	CO_2 outlet	vol. %
04-I	O_2 inlet	vol. %
04-0	O_2 outlet	vol. %
06-I	NO_x inlet	ppmv
06-0	NO_x outlet	ppmv
03	O_2 inlet air preheater*	vol. %
07	steam drum pressure	psig
08	secondary superheater outlet pressure	psig
12	flue gas inlet pressure, west	in. W.C.
14	flue gas outlet pressure	in. W.C.
15	FGD steam pressure	psig

(continued)

TABLE 5-2 (continued)

Channel No.	Parameter	Reported as
17	booster blower WP	in. W.C.
18	booster blower speed[†]	rpm
32	coal scale 11-1[*]	lb/h
34	coal scale 11-2[*]	lb/h
35	coal scale 11-3[*]	lb/h
36	coal scale 11-4[*]	lb/h
37	FGD steam flow[*]	lb/h
38	gross load[*]	MW
39	net load[*]	MW
40	FGD power consumption[†]	MW
43	combustion air temperature	°F
44	ambient air temperature	°F
45	flue gas inlet temperature, west	°F
46	boiler main steam flow	lb/h
49	secondary superheater outlet temperature	°F
52	FGD steam temperature	°F
53	flue gas outlet temperature	°F
54	flue gas outlet temperature	°F
56	flue gas inlet temperature, east	°F

[*] NIPSCO installed sensors.

[†] Davy McKee installed sensors.

The DAS control system transmitted commands every three minutes for the 29 channels of data to be sampled. Data transmitted every three minutes from the flue gas analyzers (SO_2, CO_2, O_2, NO_x) was alternated between the inlet and the outlet sample. During the six-minute interval between data sampling, the sample lines were backflushed with air and then purged with the sample gas. It was therefore necessary to synchronize the data sampling with the sample system conditioning cycle. This was accomplished with a cycle timer that automatically reset every six minutes. The SO_2 analyzer was a dual-range instrument (0-500 ppm and 0-5000 ppm). The range-changing device also had to be synchronized for correct timing of the data sampling.

Analyses of the flue gas components were determined with the instruments listed in Table 5-3.

TABLE 5-3. CONTINUOUS ANALYZERS FOR FLUE GAS

Parameter	Analyzer	Range
SO_2	Du Pont model 400 UV spectrophotometer	0-500 ppm and 0-5000 ppm
CO_2	MSA Lira model 202 NDIR	0-20 vol. %
O_2	Beckman model 742 paramagnetic	0-25 vol. %
NO_x	Beckman model 955 chemiluminescent	0-1000 ppm

All temperature measurements were by thermocouple. Where host instruments were utilized, isolation transformers were used to ensure that no feedback error occurred.

Sampling Systems

Extractive type sampling was used for sampling the flue gas at the inlet to and outlet of the absorber. The two sample lines were terminated in the instrument trailer located at some distance from the sampling positions. Thus, the sample lines were longer than desired [inlet line, about 61.0-68.6 m (200-225 feet); outlet line about 76.2-91.4 m (250-300 feet)]. However, the sampling problems that were experienced were not directly attributable to excessive line length. Long lines meant a higher number of joints, which increased the potential for sample dilution errors because of air inleakage. The samples entered the sampling system through a 5-micron in-stack filter of sintered stainless steel construction. The filters and sample probes were inserted about two feet into the ducts through flanged openings. Outside of the duct, the sample probe was terminated within a heated and insulated interface enclosure. The sample probe was connected to the sample line within this enclosure. The sample line was a flexible 3/16" ID tube made of Teflon. The line was electrically heat traced and insulated to maintain a sample temperature above the dew point.

The two lines were terminated within a heated enclosure that also housed the SO_2 analyzer. The rest of the flue gas sampling system is shown schematically on Figures 5-2 and 5-3. Negative pressure for maintaining sample flow from the process ducts at atmospheric pressure to the analyzer was provided by an air-operated aspirator for the SO_2 analyzer and by a vacuum pump for the rest of the analyzers. SO_2 and NO_x concentration were measured on a wet-gas basis. The other components were measured on a dry-gas basis. To dry the gas, the sample was passed through an air-cooled coil followed by a refrigerated condenser.

The inlet flue gas was sampled at the discharge of the boiler ID fans at the positions shown on Figure 5-4. The flue gas was taken from the two vertical sections of duct just below transition to a horizontal plenum. Samples from the two ducts were combined. Gas was initially sampled from only the east duct until it was found that the excess air was not evenly distributed between the two ID fans. Sampling

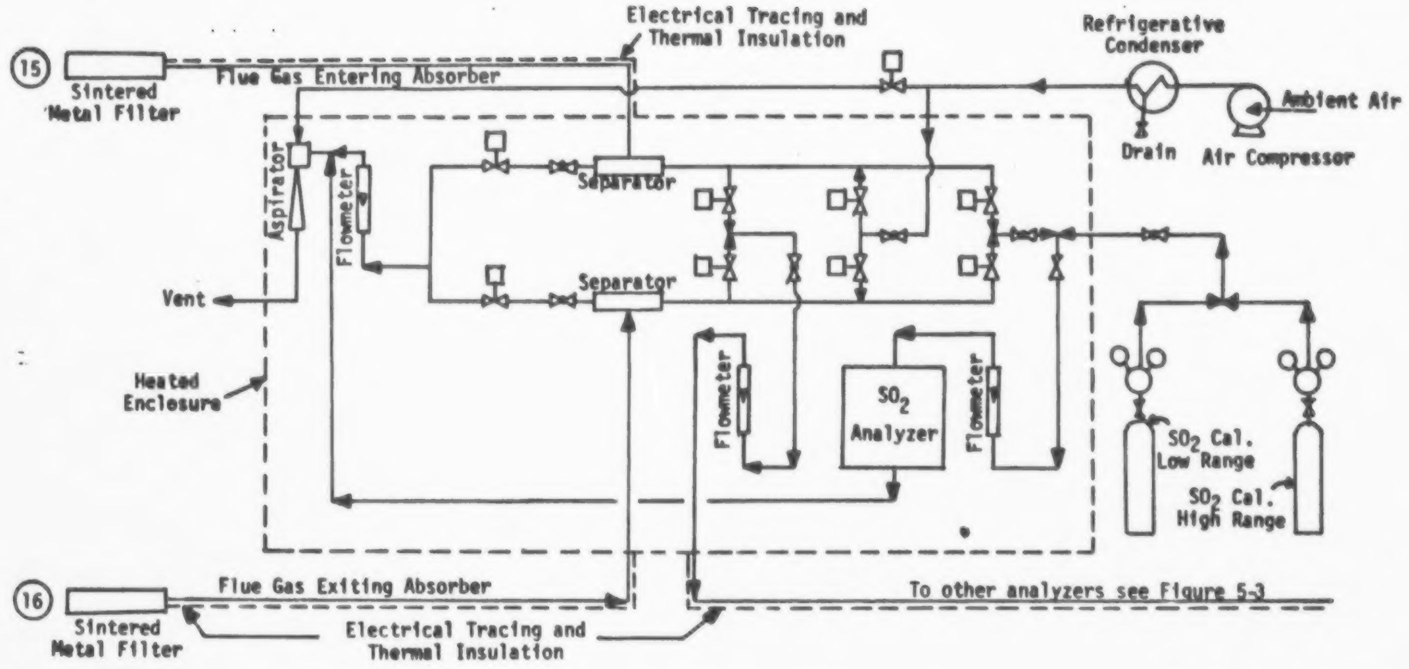

Electrical Tracing and Thermal Insulation
Refrigerative Condenser
15
Sintered Metal Filter
Flue Gas Entering Absorber
Ambient Air
Drain
Air Compressor
Aspirator
Flowmeter
Separator
Separator
Vent
Heated Enclosure
Flowmeter
SO_2 Analyzer
Flowmeter
SO_2 Cal. Low Range
SO_2 Cal. High Range
16
Sintered Metal Filter
Flue Gas Exiting Absorber
To other analyzers see Figure 5-3
Electrical Tracing and Thermal Insulation

Figure 5-2. Flue gas sampling system.

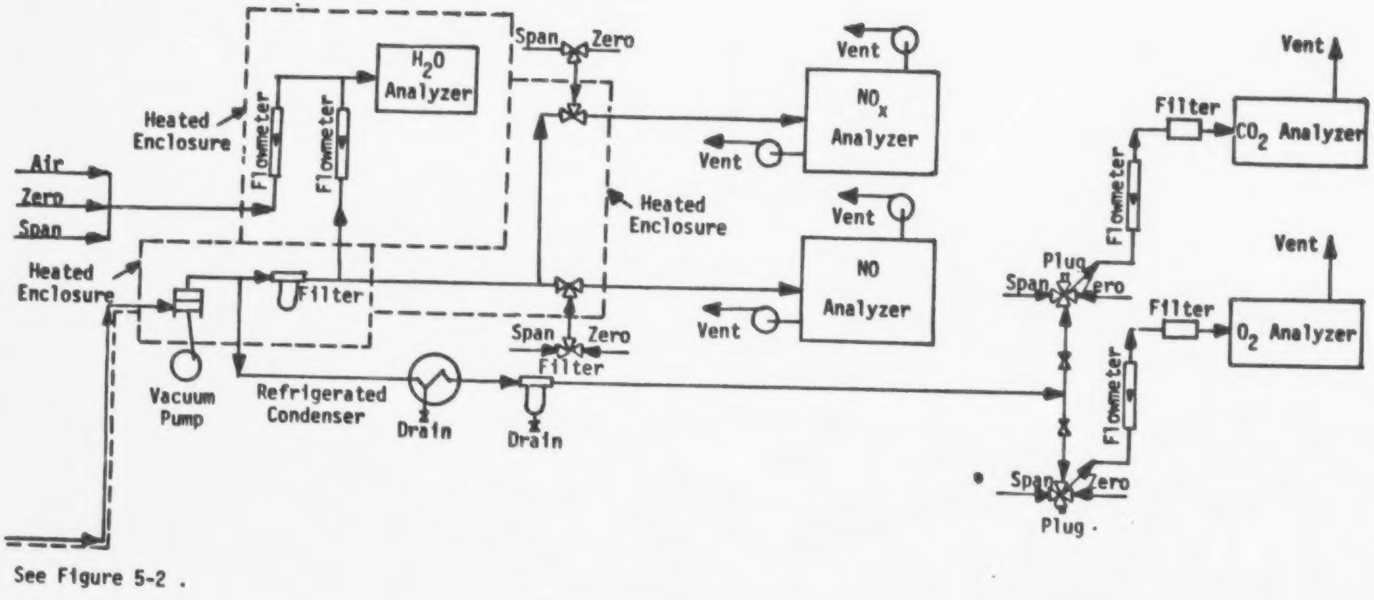

Air
Zero
Span
Heated Enclosure
Heated Enclosure
Heated Enclosure
Flowmeter
Flowmeter
H_2O Analyzer
Filter
Vacuum Pump
Refrigerated Condenser
Drain
Filter
Drain
Span
Zero
Span
Zero
Vent
NO_x Analyzer
Vent
Vent
NO Analyzer
Vent
Plug
Span
Zero
Flowmeter
Filter
CO_2 Analyzer
Vent
Span
Zero
Plug
Flowmeter
Filter
O_2 Analyzer
Vent
See Figure 5-2 .

Figure 5-3. Flue gas sampling system.

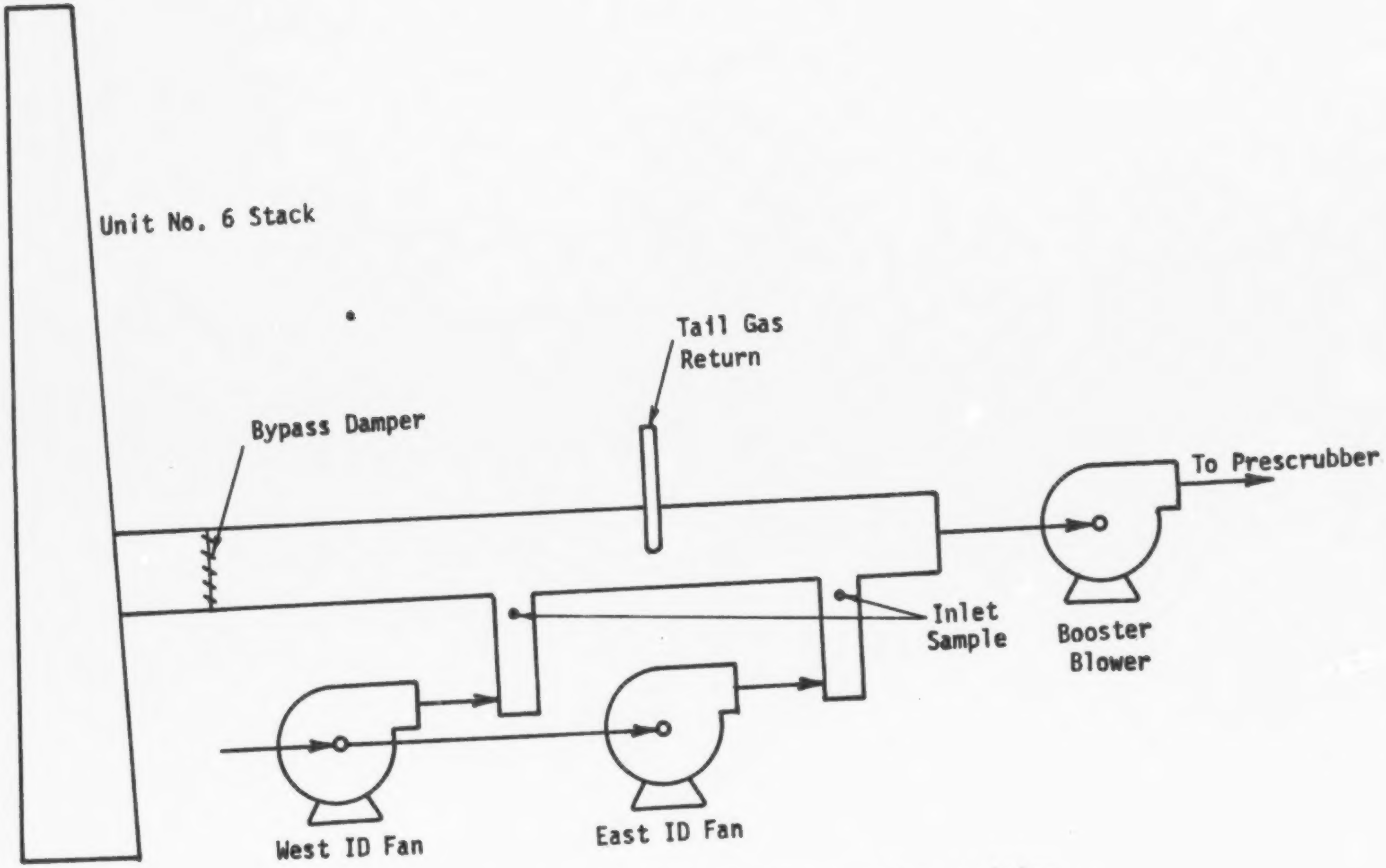

Unit No. 6 Stack
Tail Gas
Return
Bypass Damper
To Prescrubber
Inlet
Sample
Booster
Blower
West ID Fan
East ID Fan

Figure 5-4. Location of inlet flue gas sampling position.

positions on the vertical ducts were chosen to enable sampling at a location upstream of the tail gas return. The outlet flue gas was sampled at one point in the absorber stack.

Data Acquisition System (DAS)

The DAS consisted primarily of a D/A converter, a control unit, a minicomputer, a nine-track tape system and a printer (Figure 5-1). Also, a digital clock provided precise timing of all functions and a CRT terminal was available for troubleshooting. Records of the more important data channels were available from six strip chart recorders as backup in case of DAS failure. Table 5-4 lists the various components.

TABLE 5-4. DATA ACQUISITION COMPONENTS

Item	Manufacturer
Datas 901 Analog Scanner/Digitizer	Data Graphics Corp.
Dator 303/4 Data Transfer System	Data Graphics Corp.
PDP-11/05 Minicomputer	Digital Equipment Corp.
TMB11-MA Nine-Track Tape System	Digital Equipment Corp.
ASR-33 Teletypewriter	Teletype Corporation
Model 1400 Mini-Tec Data Screen Terminal	Tec., Inc.
Four-point Strip Chart Recorder (two)	Leeds and Northrop
Six-point Strip Chart Recorder (three)	Leeds and Northrop
Two-pen Strip Chart Recorder	Leeds and Northrop

The DAS had the capability of digitizing up to 64 analog voltage inputs. The control unit sent command signals that energized relays on a precisely timed cycle for sampling the analog signals. It also provided commands for sample sequence control and calibration. Status flags (on, off, calibration) were a part of the data output of each data channel.

The primary functions of the minicomputer were conversion to engineering units and averaging to obtain one-hour averages of the 10 or 20 increments of data per hour. Communication buses provided with the minicomputer were utilized for programming and commands from the printer and for directing the data output to tape storage or to be printed out. The computer software included tables of coefficients for conversion to engineering units. The coefficient values were determined from known functions or by calibration and could be changed as required. Table 5-5 lists the table values for each channel.

TABLE 5-5. CONVERSION COEFFICIENT VALUES

Channel No.	Parameter	A	B	Reported as
00-I	SO_2 inlet	5000.00	0.00000	ppmv
00-0	SO_2 outlet	500.000	0.00000	ppmv
02-I	CO_2 inlet	20.0000	0.00000	vol. %
02-0	CO_2 outlet	20.0000	0.00000	vol. %
04-I	0_2 inlet	25.0000	0.00000	vol. %
04-0	0_2 outlet	25.0000	0.00000	vol. %
06-I	NO_x inlet	1000.00	0.00000	ppmv
06-0	NO_x outlet	1000.00	0.00000	ppmv
03	0_2 inlet air preheater*	10.0000	0.00000	vol. %
07	steam drum pressure	2500.00	0.00000	psig
08	secondary superheater outlet pressure	2500.00	0.00000	psig

(continued)

TABLE 5-5 (continued)

Channel No.	Parameter	A	B	Reported as
12	flue gas inlet pressure, west	42.5530	0.00000	in. W.C.
14	flue gas outlet pressure	1.00000	0.00000	in. W.C.
15	FGD steam pressure	750.000	0.00000	psig
17	booster blower ΔP	50.0000	-5.0000	in. W.C.
18	booster blower speed[†]	5000.00	-400.00	rpm
37	FGD steam flow[*]	76000.0	0.00000	lb./h
38	gross load[*]	179.340	5.07000	MW
39	net load[*]	183.430	-3.3200	MW
40	FGD power consumption[†]	2500.00	0.00000	MW
43	combustion air temperature	1714.3	32.0000	°F
44	ambient air temperature	160.000	-40.000	°F
45	flue gas inlet temperature, west	1056.34	22.54	°F
49	secondary superheater outlet temperature	1035.33	30.3100	°F
52	FGD steam temperature	1037.24	0.00000	°F
53	flue gas outlet temperature	-	-	°F
54	flue gas outlet temperature	164.84	66.6	°F
56	flue gas inlet temperature, east	329.67	71.76	°F

$y = Ax + B$ where,

y = value in engineering units

x = analog voltage (0-1V)

[*] Denotes NIPSCO channels.

[†] Denotes Allied channels.

Although data was stored on magnetic tape, access from the tapes was not successful. The data on hard copy from the printer was relied on for data processing and analysis. The printer provided hourly averages of all data channels plus the following information:

- time, Julian day, hour/minute/minute fraction
- channel number
- units of measure
- number of samples averaged
- status (on, off, calibration)

Printed copies of three minute data were also available on command.

Supplemental Data

Data from sources other than continuous monitors were required as essential inputs to the data base. These were:

- coal
- fly ash purge
- purge solids
- natural gas consumption and heating value
- kWh readings of gross and net load
- soda ash consumption, sulfur production, and purge solids production

Samples or data were collected at set intervals. Sampling and analysis for obtaining the supplemental data are described as follows.

Coal--

Raw coal (at coal hopper discharge) was scheduled for sampling every sixth day if the FGD plant was operating. As fired coal was automatically sampled in increments from 4 of the 12 burner feed pipes. The samples were riffled and composited each day to provide a six-day composite sample. Ultimate and proximate analyses were done on each sample.

Fly ash purge--

A sample was collected every sixth day if the FGD plant was operating. A sample of lake water was also collected for a blank. Sample preservation agents were added immediately. Analytical methods are included by reference in Table 5-6.

TABLE 5-6. FLY ASH PURGE ANALYTICAL METHODS

Analysis	Method
Settleable solids	ASTM Procedure 208F
Total dissolved solids (TDS)	ASTM Procedure 208B
Total suspended solids (TSS)	ASTM Procedure 208D
Sulfate	ASTM Procedure 427C
Chloride	Modified from ASTM Procedure 408C and Orion instruation manual for 94-17 halide electrodes
Chemical oxygen demand (COD)	ASTM Procedure 508
Total Kjeldahl nitrogen	EPA Stonet No. 00625
Nitrate/nitrite nitrogen	Cadmium reduction method - EPA Storet No. 00630

Purge solids--

A total of 33 samples collected early in the first demonstration year were analyzed. Samples were collected by Allied Chemical personnel as directed by the test contractor. Analyses were for the sodium salts; sulfate, sulfite, pyrosulfite and thiosulfate; and moisture. Analytical methods recommended by Davy McKee were used.

Natural gas consumption and heating value--

Flowmeter totalizers were read daily at about 0800 and exact time was recorded. Data from two meters, process gas and incinerator gas, were collected. Daily heating values were obtained from NIPSCO.

Kilowatt-hour Readings of Gross and Net Load--

Kilowatt-hour meters were read daily at about 0800 and exact time was recorded. This data was used to validate the hourly averages obtained from the DAS.

Soda Ash Consumption, Sulfur Production, and Purge Solids Production--

Monthly totals were obtained from Allied Chemical performance summary reports.

CONTROLLED CONDITION TEST METHODS

A series of tests were conducted during the second demonstration year at specific boiler and FGD plant operating conditions for the purpose of:

1. demonstrating the load following capability of the FGD plant.
2. demonstrating turndown capability of FGD plant.
3. identifying retrofit effects relative to Baseline II boiler operating performance.
4. determining degree of sulfate formation during absorption.

Item (4) methods were described in Section 4. The objectives of Items (2) and (3) were met with pooled data from simultaneous test phases. Those test phases were interleaved with a test phase for Item (1) objectives. The various test phases were composed of the eight test sets identified in the schedule of Table 5-7. The manual sampling schedules are shown on Table 5-8.

The objectives of this test program are summarized as follows.

Test Series One

The FGD plant was in the fully integrated operating mode for the duration of the test series. The following tests were performed:

a. Observed the decreasing load-following effect on the FGD plant from 92 MW to minimum sustainable load (MSL).
b. Confirmed sustained operation by operating at MSL for four days.
c. Collected controlled test data comparable to baseline data at MSL (Test Set 3).
d. Observed the increasing load following effect on the FGD plant from MSL to 92 MW.
e. Collected controlled test data comparable to baseline data at 92 MW (Test Set 5).

Test Series Two

The FGD plant was operated with the reduction unit down. Controlled condition test data comparable to baseline data at 46 MW was collected (Test Set 7).

The load-following tests were intended to show the load following capability of the FGD process and, also, to thoroughly evaluate the potential for upsets as a result of load changes. During the load-following tests, the DAS was put into a high frequency mode to access the data at three-minute intervals and strip charts were accessed as needed to determine absorber response to load changes. Test sets

TABLE 5-7. CONTROLLED CONDITION TEST PROGRAM - SCHEDULE

Test Set No.	Day	Description	Elapsed Time, hrs	Load,MW	Manual Sampling
1	-1	Variability Profile at steady load condition.	4	92	None.
2	1	Variability Profile, load following.	2	92 to MSL	None.
3	1-5	Endurance test at MSL.	96	MSL	3-tests, Day 4-5 (see Table 5-8).
4	5	Variability Profile, load following.	2	MSL to 92	None.
5	6-7	CCTP at 92 MW.	48	92	3-tests (see Table 5-8).
6	8 or 9	Variability Profile, load following.	2	92 or 46	None.
7	8-9 or 9-10	Turndown test.	24	46	3-tests (see Table 5-8).
8	9 or 10	Variability Profile, load following.	2	46 to 92	None.

Notes:

Time allowed between Test Set No. 1 & No. 2 to develop the steady load profile to be used during Test Set No. 2.

Time allowed between Test Set No. 5 & Test Set No. 6 for shutting down reduction unit.

Load levels include equivalent steam consumed by FGD plant.

Reduction unit down during Test Sets 6, 7, 8.

MSL: minimum sustainable load.

TABLE 5-8. CONTROLLED CONDITION TEST PROGRAM - SAMPLING AND ANALYTICAL SCHEDULE

Parameters	Test Set 3	Test Set 5 No. of Samples	Test Set 7
Continuous Monitor Data	3	3	3
Boiler Data	3	3	3
Raw Coal - Ult. & Prox.	3	3	3
- Chloride		3	
AF Coal - Ult. & Prox.	3	3	3
- Screen Size	3	3	
- Trace Metals			
- Chloride		3	
Fly Ash - Combustibles/S	3	3	
- Trace Metals			
- Chloride		3	
Inlet APH - Particulate Mass			
- H_2O/Flow			
- Stationary Gases	3	3	
FGD Inlet - Particulate Mass	3	3	3
- H_2O	3	3	3
- Stationary Gases	3	3	3
- Flow	3	3	3
- SO_2/SO_3	12	12	12
- NO_x	3	3	
- Particle Size			
- Trace Metals		3	
- Chloride		3	
- H_2S			
- Mercury	3	3	
Absorber Outlet - Particulate Mass	3	3	
- H_2O	3	3	
- Stationary Gases	3	3	
- Flow	3	3	
- SO_2/SO_3	12	12	
- NO_x	3	3	
- Particle Size			
- Trace Metals		3	
- Chloride			
- H_2S			
- Mercury	3	3	
Fly Ash Purge - Chloride	3	3	
- Settleable Solids	3	3	
- TSS	3	3	
- Total Sulfur	3	3	
- TDS	3	3	
- Total Nitrogen			
- Trace Metals			

Nos. 6 and 8 were cancelled because of an unscheduled shutdown of the continuous monitoring system.

Test Procedures

Collection of continuous monitor data was not interrupted. Additional sampling and analytical procedures are described as follows in the order shown on Table 5-8.

Boiler Data--

Virtually all of the boiler control panel instruments were read every half-hour. NIPSCO Results Department personnel collected the data as directed by the test contractor.

Raw Coal--

Sample increments were collected every 15 minutes in sequence from one of the four coal hoppers. The sample increments were immediately placed in plastic bags and sealed. NIPSCO coal handlers collected the samples. At the end of each test, the sample increments were mixed by riffling and a laboratory size sample was cut from this gross sample.

As Fired Coal--

(routine procedure).

Fly Ash--

This sample was collected from the ESP hoppers. The hoppers were emptied at the start of the test. About midway into the test, grab samples were collected from each of the hoppers and mixed. NIPSCO Results Department personnel collected the samples.

Fly Ash Purge--

(routine procedure).

Flue Gas Sampling and Analysis--

The various methods are shown in Table 5-9.

TABLE 5-9. FLUE GAS SAMPLING AND ANALYTICAL METHODS

Parameter	Method
Flow rate, particulate matter, moisture	EPA Methods 1, 2, 4, 5
Stationary gases	EPA Method 3 using GC with thermal conductivity detector
Sulfur oxides	EPA Method 6 using a modified Method 5 sampling train
NO_x	EPA Method 7
Particle size	Andersen Impactor
Chloride	Sample was collected in 0.01N NaOH. Analysis was by specific ion electrode.

QUALITY CONTROL

Key elements of the quality control program were:

- Daily calibration of the primary instruments (SO_2, O_2, CO_2, analyzers), including records of drift characteristics,
- Periodic calibration of other monitors,
- Calibration gases traceable to NBS standards,
- Well-trained personnel,
- A quick-response service agreement for maintenance and repair of the minicomputer and the printer, and

- Sample line backflushing each cycle.

From the quality assurance standpoint, the test program suffered from lack of redundancy of critical components, particularly the CO_2/O_2 analyzer sampling system. It also suffered from the inability to service the D/A converter and control unit of the DAS. Also, some data was lost or invalid because the system was not attended around the clock.

Major emphasis was placed on frequent and careful calibration of the gas analyzers. A sample calibration record is shown on the following two pages. A log was kept of instrument component failures, downtimes, and uptimes. This record has been invaluable for identifying valid and invalid data.

DATA ANALYSIS METHODS

The evaluation data flow for routine monitoring is shown schematically on Figure 5-5. The various sources of data were described in a preceding subsection (The Test System).

Data Processing

Most of the data was inputted to the DAS, which provided one-hour averages for further data processing. The processing to be done was dependent on the operating status of the boiler and the FGD plant. The operating information was determined at daily meetings held at the site by the test contractor. In these meetings, the NIPSCO representative and an FGD plant operator (Allied Chemical) representative reported on operations for the preceding day as follows:

- type of coal in bunker,
- maximum and minimum boiler load,

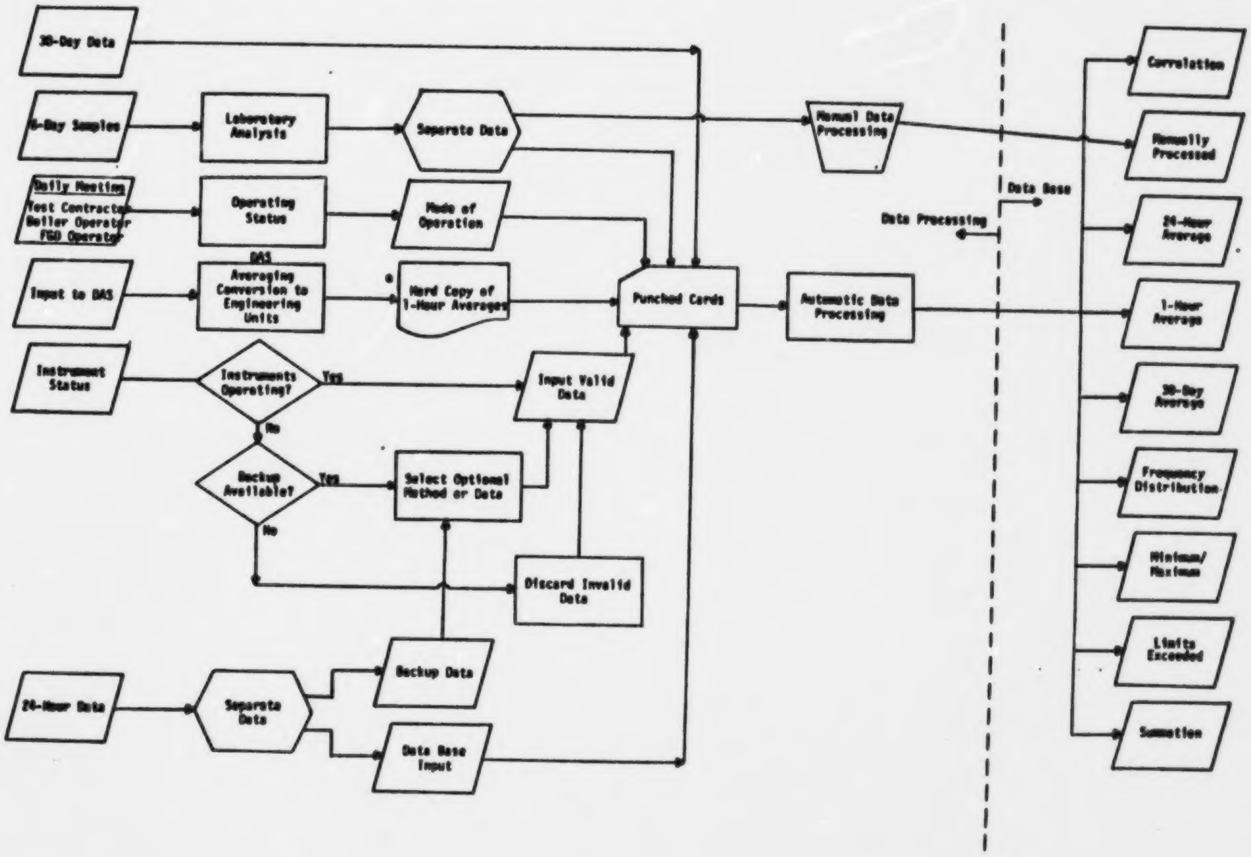

30-Day Data
-Day Samples
Daily Meeting
Test Contract
Boiler Operator
FGD Operator
Input to DAS
Instrument Status
24-Hour Data
Laboratory Analysis
Operating Status
DAS
Averaging Conversion to Engineering Units
Instruments Operating?
Yes
No
Backup Available?
Yes
No
Separate Data
Separate Data
Mode of Operation
Hard Copy of 1-Hour Averages
Select Optional Method or Data
Discard Invalid Data
Backup Data
Data Base Input
Input Valid Data
Punched Cards
Manual Data Processing
Automatic Data Processing
Data Base
Data Processing
Correlation
Manually Processed
24-Hour Average
1-Hour Average
30-Day Average
Frequency Distribution
Minimum/ Maximum
Limits Exceeded
Summation

Figure 5-5. Evaluation data flow.

- startup and shutdown times of boiler and FGD plant and their components,
- reason for outages,
- bypass damper status,
- FGD feed stream status and excursions, and
- notice of scheduled interruptions.

From this information, the time at which operating status changes occurred were tabulated and each increment of time was identified as one of five operating modes (Table 5-10).

TABLE 5-10. FGD/BOILER OPERATING MODES

Mode 0 - boiler down.

Mode 1 - boiler operating, FGD not operating

Mode 2 - integrated operation of the absorber/evaporator and reduction units without bypass of any of the flue gas.

Mode 3 - operation of absorber/evaporator with the flue gas bypass damper open. Bypassing of the flue gas may or may not occur depending on the booster blower speed setting.

Mode 4 - operation of absorber/evaporator with the reduction unit not operating. Recovered SO_2 (from evaporator overhead) is vented.

Operating mode data was used to define the extent of data analysis. As an obvious example, there would be no need for analysis of FGD plant performance during Mode 1 operation. Its major uses, however, were to determine the dependability of the boiler and of the FGD plant and its components. The items of equipment that limited plant reliability were identified from other information reported at the daily meetings.

Data accession details are listed as follows:

- The data was batch processed for each 30-day period. More frequent data processing was desirable but not feasible because of laboratory analysis lag times and 30-day inputs.
- All months were 30-days long (720 hourly records) and all references to months in this report refer to a 30-day period.
- A day encompassed the 24-hour period, 0800 to 0800.
- Each 30-day period was divided into 6-day increments for scheduling coal and fly ash purge sampling.
- The most important 24-hour data for the data base was the natural gas meter readings. Also, coal scale totalizer readings were available as backup and kilowatt-hour meter readings were used to correct the DAS gross load data for calibration error. Status of the instruments that supplied the DAS input data was the only other source of supplemental data used.
- During extended periods of DAS downtime, as occurred during November 1978, backup strip chart data was accessed to provide a partial data base.

The seriousness of any data loss due to instrument failure was, of course, dependent on which instrument failed as well as length of downtime and mode of operation. Loss of the SO_2 analyzer or plugging of a sample line was a serious loss. Oxygen or CO_2 analyzers were used to determine a dilution factor used to calculate SO_2 removal and the data from these analyzers were interchangeable for this calculation. If both instruments were down, usually due to sampling system problems, an alternate method was used to estimate the dilution factor. However, oxygen or CO_2 data were also used to calculate flue gas flow and without this data, flue gas flow, mass rate, and material balance results could not be obtained.

When all of the data from the various sources were available and the data had been screened for invalid data, data processing for a 30-day period proceeded. Software developed specifically for this program was used to make the data analyses described in the following subsection.

Data Base and Analysis

All parameters of the data base, either measured or calculated, can be located in one of the nine output areas shown on Figure 5-5. There was no need to include 24-hour or 30-day averaging of every parameter or frequency distributions, limits, summations, etc. of all parameters. Thus, the software made the necessary selections for the type of data analysis desired. Further details of the methods used and their limitations are as follows.

SO_2 Removal--

The removal performance, expressed as a percentage, may be determined as follows:

$$SO_2 \text{ Removal} = 1 - \frac{K \times SO_2 \text{ out}}{SO_2 \text{ in}}$$

$K = \frac{Q \text{ out}}{Q \text{ in}} =$ dilution factor

Q = flue gas flow (Qout and Qin at identical conditions and Qout > Qin)

The flow was calculated from coal data and excess oxygen. The method is summarized in the following subsection and is described in detail in Appendix D. Most of the dilution came from saturation of the flue gas by humidification in the prescrubber and the absorber. Flue gas flows could not be calculated without CO_2 or O_2 concentration measurements. Without either of these measurements, the optional approach was to assume K = 1.06. The flue gas water analyzer was not a reliable instrument. Thus, flue gas moisture was estimated as follows:

Inlet - Moles H_2O in air at assumed humidity + moles H_2O in coal + moles H_2O equivalent of hydrogen in coal

Outlet - Moles water calculated as a function of vapor pressure derived from temperature. Saturation was assumed. Temperature was relatively constant at about 52.8°C (127°F).

The above description of the method for determining the percentage of SO_2 removal is presented to indicate the data and estimation requirements. Actual practice was to determine SO_2 removal from the inlet and outlet mass rates of SO_2 if the flue gas flow could be calculated.

Use of Coal Analyses--

These analyses were essential for the calculation of flue gas flow rates. Analyses of six-day spot samples of raw coal were assigned to the preceding and succeeding three-day period.

Sulfur Balance--

These calculations are described in Appendix E.

Boiler Derating--

The FGD plant was an energy consumer of boiler main steam, of electrical energy, and of natural gas. Energy supplied by the boiler was a loss of generating output and affected boiler performance in various ways. The calculation scheme was as follows:

BLR.STEAM - boiler main steam, lb/h (measured or taken from Figure 5-6)

GMW - gross megawatts generated, MW/h (measured)

FGD.STEAM - steam consumed by FGD plant lb/h (corrected for temperature and pressure)

E - electrical energy consumed by FGD plant, MW-h (measured)

BHI - boiler heat input derived from coal, Btu/h (calculated)

COAL - coal rate, lb/h (measured)

HHV - heating value of coal, Btu/lb (measured)

HEAT.RATE - gross heat rate , Btu/kWh (calculated)

S2 - Btu equivalent of steam consumed by FGD plant, Btu/h (calculated)

F - fraction of boiler heat input consumed by FGD plant as steam (calculated)

MW.EQUIV - gross megawatts equivalent to steam consumed by FGD plant, MW (calculated)

LOAD.POT - generating potential, MW

MW.EQUIV (total) - gross megawatts equivalent to total energy from boiler consumed by FGD plant, MW (calculated)

DERATE - boiler derating, % (calculated)

Calculations -

$$\text{BHI} = \text{COAL} \times \text{HHV}$$

$$\text{HEAT.RATE} = \frac{\text{BHI} \times 10^3}{\text{GMW}}$$

$$S2 = \frac{GMW \times HEAT.RATE \times FGD.STEAM \times 10^3}{BLR.STEAM - (0.9158 \times FGD.STEAM)}$$

$$F = \frac{S2}{BHI}$$

$$MW.EQUIV = \frac{S2 \times GMW \times 10^3}{BHI - S2}$$

$$LOAD.POT = GMW + MW.EQUIV$$

$$MW.EQUIV\ (total) = MW.EQUIV + E$$

$$DERATE = \frac{MW.EQUIV\ (total)}{115.6} \times 100$$

Boiler main steam was not transmitted to the DAS until late in the program. As backup, a correlation of steam rate with boiler load was used (Figure 5-6).

Load Correction--

There was a bias error in the DAS output as indicated on Figure 5-7. Corrections were applied to the gross load data, using the regression equation,

$$X = 1.07099Y - 8.48013$$

where X = kWh meter, MW

Y = DAS output, MW

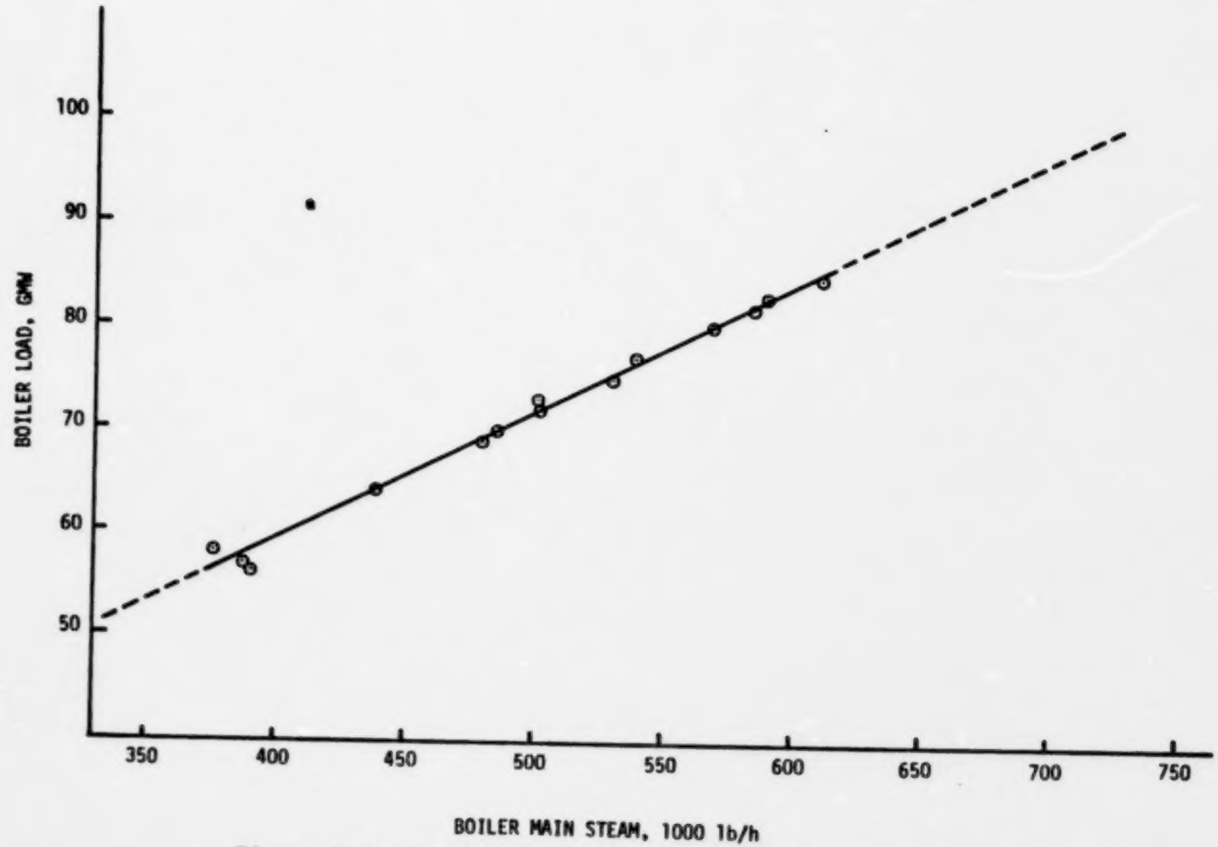


Figure 5-6. Boiler load as a function of boiler main steam rate.

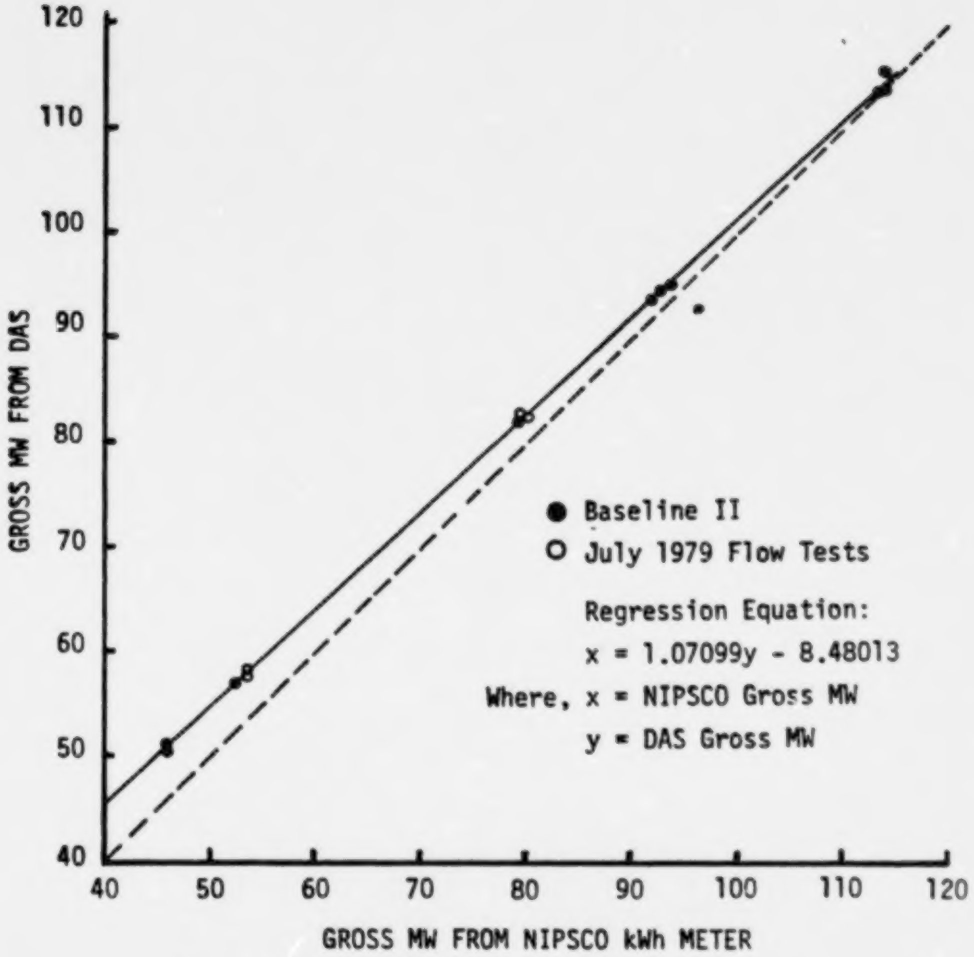


Figure 5-7. Correction factor for DAS gross megawatts.

Soda Ash Consumption--

The tonnage consumed was reported once a month. Consumption in tons per day was determined as follows:

$$\text{ton/day} = \frac{\text{ton/month}}{\text{hours (mode 2 + mode 3 + mode 4)}} \times 24$$

Sulfur Production--

The tonnage was reported once a month in long tons. Daily sulfur production was determined as follows:

$$\text{lb/day} = \frac{\text{long tons/month} \times 2240}{\text{hours (mode 2 + mode 3)}} \times 24$$

FLUE GAS FLOW RATE

The available techniques for continuous measurement of gas flow rate in large ducts are generally unreliable unless long lengths of straight duct are available. Space limitations and economics dictated close coupling along the inlet duct system and swirling action in the absorber stack precluded any accurate flow measurement at the outlet. To provide the necessary flue gas flow measurements, a calculation procedure was developed. It was validated with actual flow measurements made during acceptance and baseline testing and controlled condition tests. Validation of the method is reported in detail in the Baseline II report. The calculation procedure is summarized in the following paragraphs. In Appendix D, the detailed calculation procedure and curves for quick estimates are given.

The flue gas flow calculation is based on a carbon balance across the boiler or across the boiler and the FGD plant. That is, knowing the amount of carbon fed to the boiler as coal and knowing the concentration of CO_2 in the flue gas, the volume of flue gas can be calculated as follows:

$$\text{DRY FLUE GAS FLOW} = \frac{\text{Carbon in coal, mols/h}}{CO_2 \text{ in the flue gas, volume \% dry}}$$

It can be assumed that other sources and sinks for carbon are small enough to be either ignored or estimated with negligible error. The calculation method minimizes these errors, to the extent possible, as shown (Table 5-11).

TABLE 5-11. ASSUMED VALUES OF CARBON SOURCES AND SINKS

Type		Assumptions
CO_2 in Entering Air	Source	Negligible.
Unburned Carbon in Ash	Sink	0.5% by weight.
CO Due to Incomplete Combustion	Sink	No incomplete combustion.
CO_2 in Incinerator Tail Gas	Source	Proportioned according to quantity of SO_2 in flue gas; roughly 0.5% of flue gas volume.
CO_2 Liberated from Soda Ash Makeup	Source	Included but only 0.10% of total CO_2.

For a precise calculation the total data requirements are as follows:

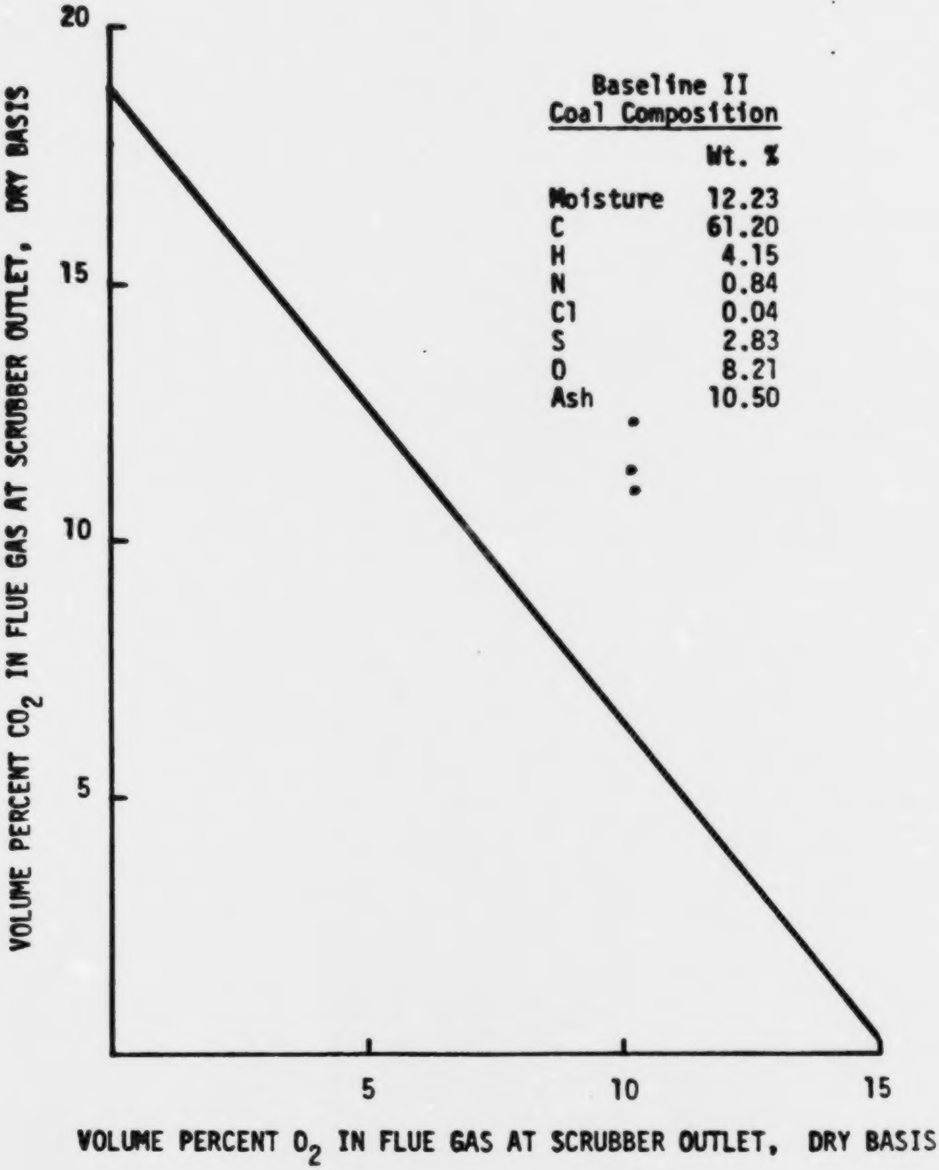
Item	Availability
1. Coal rate	Measured. One hour averages available from data acquisition system (DAS). Also available from NIPSCO coal scale totalizers.

2.	Coal analysis - % Moisture - % Carbon - % Hydrogen - % Nitrogen - % Sulfur - % Oxygen - % Ash	Measured. Raw coal analysis available no more often than every six days.
3.	Percent carbon in coal not oxidized.	Estimated at 0.5% by weight
4.	Volume % CO_2 in flue gas (dry basis)	Measured. One hour averages available from DAS.
5.	SO_2 absorption efficiency	Estimated at 90%.
6.	Ambient air temperature	Measured or estimated. One hour averages available from DAS.
7.	Ambient air saturation	Estimated.
8.	Scrubber inlet temperature	Measured. One hour averages available from DAS.
9.	NO_x formation	Ignored.
10.	CO formation	Ignored.
11.	SO_3 formation	Ignored.
12.	Unoxidized sulfur	Ignored.
13.	Natural gas feed rate to FGD incinerator and SO_2 reduction, and composition of natural gas.	Feed rate measured, or estimated from process flow sheets. Composition from process flow sheets.

Where assumed values have to be used, a sensitivity analysis showed that relatively small errors are generated.[9] Furthermore, it was shown that the results calculated by this method agree reasonably well with measured flue gas volumes.

The stepwise calculation procedure is shown in Appendix D. The calculation procedure describes a method for determining the flue gas flow from the CO_2 concentration of the flue gas. However, if CO_2 data are not available, an alternative is to calculate the flue gas flow from the O_2 concentration. Either the O_2 concentration or the CO_2 concentration may be used as the independent variable because, on a dry flue gas basis, setting the O_2 concentration also sets the CO_2 concentration (Figure 5-8).

Another potential material balance is based on SO_2. The wet inlet flue gas flow can be calculated from the molar flow of SO_2, and the flue gas SO_2 concentration on a wet basis. However, since sulfur is present in the coal at only 3% by weight compared to 60% for carbon, a sulfur balance would vary more with changes in coal composition.



	Wt. %
Moisture	12.23
C	61.20
H	4.15
N	0.84
Cl	0.04
S	2.83
O	8.21
Ash	10.50

Equation: $Y = 18.63 - 0.8905X$

Where, X = Volume % O_2 in flue gas at scrubber outlet

Y = Volume % CO_2 in flue gas at scrubber outlet

Figure 5-8. CO_2 concentration of flue gas as a function of the oxygen concentration (Baseline II Coal).

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APPENDIX A
BACKUP DATA

TABLE A-1. SO_2 REMOVAL PERFORMANCE

	Oct. 1978	Nov.	Dec.	Jan. 1979	Feb.	Mar.	Apr.	May.	Jun.	Jul.	Aug.	Sept.	Oct.
SO_2 Removal, %	91	89	90	(1)	(1)	90	88	89	90	90	89	90	93
Avg. SO_2 feed, kg/s (lb/h)	.6543 (5193)	.5521 (4382)	.5582 (4430)	(1) (1)	(1) (1)	.5960 (4730)	.6174 (4900)	.6370 (5056)	.5162 (4097)	.5551 (4406)	.5541 (4398)	.6281 (4985)	.5728 (4546)
Max. Sustained SO_2 feed(2), kg/s (lb/h)	.6888 (5467)	.6363 (5050)	.6048 (4800)	(1) (1)	(1) (1)	.6152 (4883)	.6867 (5450)	.6914 (5487)	.6309 (5007)	.7497 (5950)	(3) (3)	.7315 (5806)	(3) (3)
Determined from	coal(4)	coal	coal	-	-	coal	coal	analysis(5)			coal	analysis	coal
Avg. SO_2 emitted, kg/s (lb/h)	.059 (467)	.061 (482)	.056 (443)	(1) (1)	(1) (1)	.060 (473)	.074 (588)	.070 (556)	.052 (410)	.056 (441)	.061 (484)	.063 (498)	.040 (318)
Avg. SO_2 removed, kg/s (lb/h)	.5955 (4726)	.4914 (3900)	.5024 (3987)	(1) (1)	(1) (1)	.5364 (4257)	.5433 (4312)	.5670 (4500)	.4646 (3687)	.4996 (3965)	.4932 (3914)	.5654 (4487)	.5327 (4228)
Inlet SO_2 concentration, ppmv, wet	2148	2239	2174	2180	(1)	2155	2045	2203	1873	2087	1801	2113	2148
Load range, gross MWe(6)	66-87	65-82	67-82	74-80	(1)	80-86	56-91	62-92	65-80	51-84	67-87	56-85	61-81
Avg. Coal rate, kg/s (lb/h)	10.4 (82400)	(1) (1)	9.48 (75200)	(1) (1)	(1) (1)	10.5 (83600)	9.82 (77900)	10.6 (84300)	9.35 (74200)	8.86 (70300)	9.78 (77600)	9.51 (75500)	9.32 (74000)

(1) Not determined - insufficient data.
(2) Sustained for at least 24 hours.
(3) Not determined - low relative to other months.
(4) Sulfur in coal corrected for about 2% loss in ash.
(5) SO_2 in, ppmv and calculated flow rates.
(6) 95% of data.

TABLE A-2. SO_2 REMOVAL - 24-HOUR AVERAGE

Julian Day	Calendar Day	% Removal
293	10/20/78	90
295	10/22/78	91
296	10/23/78	92
297	10/24/78	90
298	10/25/78	91
342	12/08/78	89
343	12/09/78	88
344	12/10/78	85
356	12/22/78	90
357	12/23/78	91
358	12/24/78	91
359	12/25/78	90
360	12/26/78	90
361	12/27/78	91
362	12/28/78	91
363	12/29/78	91
364	12/30/78	91
365	12/31/78	90
366	01/01/79	92
92	04/02/79	83
93	04/03/79	86
94	04/04/79	88
95	04/05/79	90
96	04/06/79	90
97	04/07/79	89
98	04/08/79	91
100	04/10/79	89
101	04/11/79	90
102	04/12/79	87
103	04/13/79	89
104	04/14/79	89
105	04/15/79	89
106	04/16/79	89
107	04/17/79	88
108	04/18/79	89
109	04/19/79	88
110	04/20/79	88
111	04/21/79	89
114	04/24/79	87
115	04/25/79	88
116	04/26/79	88
117	04/27/79	90

(continued)

TABLE A-2 (continued)

Julian Day	Calendar Day	% Removal
127	05/07/79	89
128	05/08/79	89
129	05/09/79	88
130	05/10/79	89
131	05/11/79	89
132	05/12/79	88
135	05/15/79	89
136	05/16/79	89
137	05/17/79	90
138	05/18/79	89
139	05/19/79	90
141	05/21/79	89
142	05/22/79	90
145	05/25/79	90
146	05/26/79	91
147	05/27/79	90
148	05/28/79	90
149	05/29/79	90
150	05/30/79	89
151	05/31/79	89
159	06/08/79	91
160	06/09/79	90
161	06/10/79	90
162	06/11/79	90
163	06/12/79	90
164	06/13/79	90
165	06/14/79	89
166	06/15/79	90
170	06/19/79	86
171	06/20/79	90
172	06/21/79	92
173	06/22/79	91
178	06/27/79	89
179	06/28/79	90
180	06/29/79	90
181	06/30/79	88
182	07/01/79	89
183	07/02/79	90
184	07/03/79	90
185	07/04/79	91
186	07/05/79	90
187	07/06/79	90

(continued)

TABLE A-2 (continued)

Julian Day	Calendar Day	% Removal
188	07/07/79	90
189	07/08/79	90
190	07/09/79	91
191	07/10/79	92
192	07/11/79	90
193	07/12/79	90
194	07/13/79	90
198	07/17/79	90
199	07/18/79	88
200	07/19/79	91
201	07/20/79	91
202	07/21/79	89
203	07/22/79	90
204	07/23/79	89
205	07/24/79	89
206	07/25/79	89
207	07/26/79	89
221	08/09/79	90
235	08/23/79	81
236	08/24/79	91
240	08/28/79	91
241	08/29/79	90
242	08/30/79	89
243	08/31/79	88
244	09/01/79	90
245	09/02/79	92
249	09/06/79	94
255	09/12/79	90
257	09/14/79	91
258	09/15/79	92
259	09/16/79	89
260	09/17/79	88
261	09/18/79	87
262	09/19/79	89
263	09/20/79	89
265	09/22/79	89
266	09/23/79	89
267	09/24/79	90
268	09/25/79	89
269	09/26/79	79
270	09/27/79	89
271	09/28/79	91

(continued)

TABLE A-2 (continued)

Julian Day	Calendar Day	% Removal
272	09/29/79	90
276	10/03/79	93
277	10/04/79	93
278	10/05/79	93
279	10/06/79	92
280	10/07/79	93
281	10/08/79	94
282	10/09/79	93
283	10/10/79	93
285	10/12/79	94
286	10/13/79	93
288	10/15/79	91
289	10/16/79	92
290	10/17/79	91
291	10/18/79	90

TABLE A-3. FGD PLANT OPERATING SUMMARY

Month		Boiler operation (hours)	Total FGD full operation (hours)	Full FGD absorber operation (hours)	operation with by-pass open (hours)
Oct.	1978	691	324	386	62
Nov.		717	705	713	1
Dec.		659	434	504	45
Jan.	1979	720	157	190	34
Feb.		719	8	314	0
March		720	264	285	16
April		719	511	592	11
May		677	498	562	0
June		628	316	582	115
July		717	538	695	0
Aug.		564	404	511	26
Sept.		707	501	498	19
Oct.		591	419	500	9
Total time		8829	5079	6432	338
% of total time		96.5	55.5	70.3	3.7
% of boiler operation		100.0	57.5	72.9	3.8

Note: Total time, 9145 hours after adjustment for interruption for testing.

TABLE A-4. FGD PLANT VIABILITY INDICES

Month	Hrs. FGD available	Hrs. in period	% Availability $\frac{\text{hrs. FGD available}}{\text{hrs. in period}}$	Hrs. FGD operated	Hrs. FGD called upon	% Reliability $\frac{\text{hrs. FGD operated}}{\text{hrs. FGD called upon}}$	Boiler operating hrs.	% Operability $\frac{\text{hrs. FGD operated}}{\text{boiler operating hr.}}$	% Utilization $\frac{\text{hrs. FGD operated}}{\text{hrs. in period}}$	Absorber Availability %
Oct. 1978	577	729	69	324[1]	505	64	691	42	44	74
Nov.	712	720	99	705	714	99	717	98	98	100
Dec.	439	720	61	434	657	66	659	66	60	100
Jan. 1979	157	720	22	157	720	22	720	22	22	100
Feb.	58	720	8	8	718	1	719	1	1	100
March	264	720	37	264	720	37	720	37	37	100
April	515	719	72	511	715	71	719	71	71	100
May	584	720	81	498	670	74	677	74	69	100
June	625	720	87	316	459	69	628	50	44	99
July	584	720	81	538	680	79	717	75	75	100
Aug.	699	720	97	404	536	75	564	72	56	100
Sept.	528	720	73	501	704	71	707	71	70	100
Oct.	549	721	76	419	591	71	591	71	58	100

[1] Includes 15 hrs. at high load with by-pass open.

TABLE A-5. FGD PLANT LIMITATIONS

Date	COMPONENT CAUSING FGD PLANT OUTAGE								
	Mitchell no. 11 boiler	Duct	Absorber	Purge removal	Evaporator-crystallizer	SO_2 reduction	Pumps	Blowers and compressors	Instruments and controls
	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	
1978									
Oct. 5 - Oct. 6	20.7								
Oct. 13 - Oct. 16								96.0	
Oct. 18								7.3	
Oct. 26								23.1	
Nov. 7		4.7							
Nov. 24						1.1			
Nov. 28 - Nov. 29	2.7								
Dec. 5 - Dec. 6	2.8				2.2				
Dec. 7 - Dec. 8						6.7			
Dec. 11								0.2	
Dec. 12 - Dec. 19					190.5				
Dec. 20 - Dec. 22						56.3			
1979									
Jan. 4							16.0		33.5
Jan. 11 - Feb. 15						850.0			
Feb. 24	1.2								
Feb. 25 - Feb. 27					28.4	277.9	68.3[1]		
Feb. 28								0.5	
Mar. 3 - Mar. 5						52.5			
Mar. 9 - Mar. 25						2.5	382.5		
Apr. 3					1.7	8.4			
Apr. 22									
Apr. 23					25.5				19.9
Apr. 27 - May 5						18.6	204.0		

[1] Due to lack of steam.

TABLE A-5 (continued)

Date	COMPONENT CAUSING FGD PLANT OUTAGE								
	Mitchell no. 11 boiler	Duct	Absorber	Purge removal	Evaporator-crystallizer	SO_2 reduction	Pumps	Blowers and compressors	Instruments and controls
	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	FGD plant outage hours	
May 12 - May 14	47.4								
May 15					5.0				
May 26						0.2			
May 31 - June 5	122.4								
June 21			6.5						
June 25 - June 26		27.8							
July 6								1.8	
July 10									
July 14 - July 17						76.4			
July 16								6.9	
July 30						39.1		11.9	
July 30-31	5.6								
Aug. 14 - Aug. 21	161.1								
Aug. 25 - Aug. 27	48.4								
Aug. 30 - Aug. 31	12.4								
Sept. 4 - Sept. 5								38.8	
Sept. 7 - Sept. 10								79.9	
Sept. 21								35.1	
Sept. 24	2.9								
Oct. 1								8.8	
Oct. 8		2.9				1.3			
Oct. 10 - Oct. 12								48.7	
Oct. 19 - Oct. 20						30.8			
Oct. 19 - Oct. 23								79.2	
Oct. 23 - Oct. 29	128.6								
Total	556.2	35.4	6.5		253.3	1421.8	670.8	438.2	53.4

TABLE A-6. FGD TIME DOWN BEFORE AND AFTER BOILER OUTAGES

FGD plant component	Before boiler outage (hours)			After boiler outage (hours)		
	avg.	range	std. dev.	avg.	range	std. dev.
Steam supply	1.5	0.0-11.2	3.6	3.3	0.2-12.8	3.8
Absorber	3.7	0.0-20.6	6.5	12.0	0.0-68.3	21.0
Evaporator	3.7	0.0-20.6	6.5	12.0	0.0-58.3	21.0
Reduction	7.5	0.0-28.4	10.3	35.6	0.2-136.2	40.6
By-pass	6.7	0.0-28.4	10.4	37.4	0.0-136.2	45.8
Full FGD operation	7.5	0.0-28.4	10.8	41.2	0.2-136.2	44.1

TABLE A-7. HOURS WITH BYPASS OPEN

Period	Boiler operating hours	Hours bypass open	% of boiler hours
Oct.	691	61.7	8.9
Nov.	717	1.2	0.2
Dec.	639	44.9	6.8
Jan.	720	33.5	4.7
Feb.	719	0.3	Neg.
March	720	15.9	2.2
April	719	11.3	1.6
May	677	0.3	Neg.
June	628	115.2	18.3
July	717	0.2	Neg.
Aug.	564	25.5	4.5
Sept.	707	19.3	2.7
Oct.	591	8.8	1.5
Total	8829	338.1	3.8

TABLE A-8. BOILER AND FGD PLANT OPERATING HISTORY

Event	Period	Boiler operated	FGD plant full operation
1. Boiler down for planned annual maintenance until 10/4. Boiler went down again on 10/5 because of valve leak. After the boiler came up on 10/6, the FGD plant remained down until 10/13 because of the Baseline II tests. From 10/13 to 10/16 the FGD plant was down so that the booster fan could be balanced.	1978 09/29-10/16	263	0
2. Full FGD plant operation interrupted by three FGD plant outages. The first outage was caused by the booster fan oil cooler; the second by the booster fan turbine governor; and the third by a duct leak at the booster fan exit.	10/16-11/17	527	449
3. Full FGD plant operation except for reduction area being down one hour to repair a coalescer valve.	11/07-11/28	514	513

(continued)

TABLE A-8 (continued)

Event	Period	Boiler operated	FGD plant full operation
	1978 (continued)		
4. Periods of full FGD plant operation were interrupted by boiler and reduction area outages. The boiler went down on 11/28 to replace the modulating servo on the intercept valve. On 12/5 the boiler went down because of high silica in the feedwater. Two of the reduction outages were caused by fluctuating steam pressure. In addition, reduction was down to repair the SO_2 superheater and a leak in B generator.	11/28-12/11	294	277
5. The FGD plant went down three times during this period because of an accidental trip of the booster fan; to clean the evaporator tubes; and to repair reduction valves. Between outages, there was some FGD plant operation with the by-pass open.	12/11-12/22	211	0
	1978/1979		
6. Full FGD plant operation except for two periods with the by-pass open. During the second by-pass open period, lines to the inlet duct pressure transmitter were repaired.	12/22-01/10	447	410

(continued)

TABLE A-8 (continued)

Event	Period	Boiler operated	FGD plant full operation
	1978/79 (continued)		
7. FGD plant down to repack evaporator circulating pump and to retube the sulfur condenser. Absorber-evaporator up on 2/15, but reduction still down because of water in the Claus converter.	01/10-02/22	1038	0
8. Full operation of FGD plant was limited by boiler, reduction area and pump problems. A vacuum pump tripped off on 2/24 and shut the boiler down. The three FGD plant outages were caused by the following: loss of seal water to the evaporator circulating pump packing, malfunction of the oil pressure switch on the booster fan, and leaks in the Claus converter expansion joints. Outages of the reduction area alone were caused by leaks in the Claus converter expansion joints, SO_2 superheater leaks, and steam leaks in the #3 reflux condenser.	02/22-03/09	355	84
9. FGD plant down so that evaporator circulating pump could be overhauled and realigned.	03/09-03/25	386	0

(continued)

TABLE A-8 (continued)

Event	Period	Boiler operated	FGD plant full operation
10. Full FGD plant operation with short interruptions due to reduction area problems. The FGD plant went down once because of a leak in the SO_2 superheater. Reduction area went down three times because of a plugged line to the tail gas incinerator, and once because of low steam pressure.	03/25-04/09	364	347
11. Full FGD plant operation	04/09-04/19	242	242
12. Reduction problems prevented sustained full FGD plant operation. The reduction area was shut down four times because the tail gas incinerator wasn't operating properly. The FGD plant went down once so the valves on the SO_2 compressor could be repaired. In addition, on 4/26 the by-pass damper was opened for 11 hours because of high duct pressure.	04/19-04/27	186	110
13. The FGD plant went down so that the motor shaft on the evaporator circulating pump could be repaired.	04/27-05/07	243	0
14. The boiler went down on 5/12 so that boiler tube leaks could be repaired. The FGD plant went down for three hours on 5/15 to permit replacement of the SO_2 superheater.	05/07-05/15	160	119

(continued)

TABLE A-8 (continued)

Event	Period	Boiler operated	FGD plant full operation
15. Full FGD plant operation interrupted only by a loss of natural gas for 10 minutes which shut down reduction area.	05/15-05/31	379	379
16. The boiler went down on 5/31 for less than an hour because of a yard problem, and then went down again on 5/31 so that the turbine stop valve could be repaired. During the FGD plant startup, the by-pass damper was open for 29 hrs because of problems with the west ID Fan.	05/31-06/08	98	0
17. Pipe and duct leaks, a lack of utilities, and flue gas supply problems prevented full FGD plant operation. The FGD unit was by-passed for four short periods because the air to control the dampers was lost, and for 82 hours because of a problem with the west ID fan. The entire FGD plant went down for 34 hours so that leaks in the inlet flue gas duct could be repaired. The absorber went down once because of a lack of makeup water caused by obstruction in a valve. Reduction area and the evaporator went down twice due to a lack of steam. Because of low inlet SO_2, reduction went down for 51 hours on 6/23. In addition, reduction area was down for 28 hours so a leaking expansion joint could be repaired.	06/08-06/30	525	297

(continued)

TABLE A-8 (continued)

Event	Period	Boiler operated	FGD plant full operation
18. Full FGD plant operation	06/30-07/06	140	140
19. A booster fan reduction gear problem shut down the absorber once, and the entire FGD plant once. The FGD plant also went down once to permit the repair of valves in the reduction area.	07/06-07/17	266	181
20. Full FGD plant operation.	07/17-07/27	236	236
21. The boiler went down for three hours on 7/30 because of a switching problem. The FGD plant went down once to permit inspection of the booster fan coupling. Reduction went down for the load following test, and remained down after the testing because of a restriction in the mixed gas cooler.	07/27-08/01	124	0
22. Full FGD plant operation.	08/01-08/14	320	320
23. This period is characterized by boiler outages. The boiler went down three times: (1) on 8/14 for 138 hours because of a short in the auxiliary breaker and damaged reheat pipe hangers, (2) on 8/26 for 18 hours to permit ESP repairs; and (3) on 8/30 for 11 hours due to a high voltage generator trip.	08/14-08/31	233	92

(continued)

TABLE A-8 (continued)

Event	Period	Boiler operated	FGD plant full operation
24. Booster fan problems shut down the FGD plant twice. The first shut down occurred when the booster fan tripped off due to high vibration; the second when the linkage to the booster fan governor broke.	08/31-09/11	260	104
25. Full FGD operation, except for 6 hours when the by-pass opened automatically on duct, pressure due to a problem with the booster fan.	09/11-09/21	236	230
26. The boiler went down for 3 hours on 9/24 because of a 130 Kva yard problem. Also, during this period, the FGD plant went down for 9 hours on 9/21 because of a steam leak at the booster fan turbine.	09/21-09/24	74	42
27. Full FGD plant operation.	09/24-10/01	166	166
28. Full FGD plant operation was limited by booster fan and reduction area problems. The booster fan was shut down 4 times to work on the booster fan governor, and once by low oil pressure at the booster fan turbine. In reduction area, the converter was raked and a valve leak repaired.	10/01-10/12	270	208

TABLE A-8 (continued)

Event	Period	Boiler operated	FGD plant full operation
29. Full FGD plant operation.	10/12-10/19	162	162
30. Booster fan and reduction area prevented full operation of the FGD during the first half of this period. Reduction area was down to change a coalescer mist pad and due to a reactor hot spot. During the second half, the boiler was down because of broken turbine blades. Remained down for scheduled maintenance.	10/19-10/29	110	0

TABLE A-9 ABSORBER TURNDOWN TEST

Test No.	7	8	9
Date	07/28/79	07/28/79	07/28/79
Time	0900-1056	1200-1410	1500-1658
Gross Load, MWe	44	43	44
Generating Potential, MW	51	49	50
CO_2 in, vol.% dry	10.76	10.02	10.81
O_2 in, vol.% dry	8.77	8.32	9.66
H_2O in, vol.% dry	6.19	6.44	7.48
Flue gas flow measured, m^3/s (acfm)	105.95 (224500)	110.91 (235000)	111.95 (235100)
Flue gas flow calculated, m^3/s (acfm)	101.89 (215900)	109.92 (232900)	111.52 (236300)
SO_2 in ppmv wet	1709	2143	2156
Coal rate, kg/s (lb/h)	6.11 (48500)	6.11 (48500)	6.06 (48100)
Sulfur in coal, wt %	3.28	3.28	3.28
Sulfur in coal, kg/s (lb/h)	0.1992 (1581)	0.2005 (1591)	0.1988 (1578)

TABLE A-10. MAXIMUM SUSTAINABLE LOAD TEST

Date	Test No. 4 7/25/79		Test No. 5 7/26/79		Test No. 6 7/26/79	
Hours from 1 to 65 (65 hour test)	22-27		44-49		51-55	
Gross load, MWe	79.60		79.80		79.33	
Generating potential, MW	89.15		89.53		89.12	
Coal rate, kg/s (lb/h)	9.74 (77,300)		10.14 (80,480)		10.18 (80,760)	
Sulfur in coal, wt.%	3.60		2.82		3.20	
Boiler steam, kg/s (lb/h)	72.4 (575,000)		73.1 (580,000)		70.6 (560,000)	
SO_2 removal, %	89.1		90.2		-	
FGD steam consumption, kg/s (lb/h)	6.7840 (53842)		7.6008 (60325)		7.6418 (60650)	
Boiler heat input, J/s x 10^6, (10^6 Btu/h)	248 (847)		257 (877)		259 (883)	
	Continuous	Manual	Continuous	Manual	Continuous	Manual
Flue gas flow, m^3/s (acfm)	158.162 (335,127)	157.467 (333,654)	151.773 (321,589)	159.856 (338,715)	147.812 (313,196)	157.292 (333,284)
SO_2 in ppmv wet	2047	2423[2]	2080	2131[2]	2023	2229[2]
SO_2 in, kg/s (lb/h)[1]	.5963 (4733)	.7060 (5603)	.6152 (4883)	.6304 (5003)	.5528 (4387)	.6488 (5149)
O_2 in flue gas, vol.% dry	7.02	6.08	6.70	6.57	7.00	7.06
H_2O in flue gas, vol.%	(3)	6.57	(3)	6.76	(3)	6.98

[1] Using calculated flow rates.
[2] Includes SO_2 in tail gas.
[3] Not determined.

TABLE A-11. VARIATIONS OF PROCESS VARIABLES DURING LOAD FOLLOWING TESTS

Test	Time	Avg. Gross MW	No. Data Points	% SO_2 Removal		SO_2 Removed/ Gross MW				Steam/SO_2 Removed, (lb/lb)^2 (kg/kg)		Steam/ Gross MW			
				Mean	Std. Dev.	Mean (lb)	Mean kg	Std. Dev. (lb)	Std. Dev. kg	Mean	Std. Dev.	(lb)	kg	(lb)	kg
Steady 80 MW load 07/18/79	9:02-11:44	82.7	53	89.5	0.3	(54.8)	24.8	(0.9)	0.4	8.5	0.3	(517)	234	(15)	6.8
Steady Min. sustainable load (MSL) 7/24/79	12:04-12:56	57.3	18	89.4	0.2	(53.8)	24.4	(0.6)	0.3	-	-	-	-	-	-
Load following 80-57 MW 7/19/79 Load change	10:52-11:32	69.2	13	88.6	2.3	(54.7)	24.8	(0.6)	0.3	9.7	1.1	(601)	273	(68)	31
After load change	11:34-13:02	61.0	28	90.7	0.3	(57.2)	25.9	(0.6)	0.3	10.4	0.4	(655)	297	(25)	11
Last hour of test	12:02-13:02	61.0	20	90.8	0.3	(57.1)	25.9	(0.6)	0.3	10.4	0.4	(651)	295	(23)	10
Load following 57-80 MW 7/24/79 Load change	13:02-13:32	68.2	11	89.1	0.2	(56.1)	25.4	(2.4)	1.1	-	-	-	-	-	-
After load change	13:34-15:56	82.4	46	88.7	0.5	(54.1)	24.5	(0.9)	0.4	-	-	-	-	-	-
Last hour of test	14:56-15:56	82.4	20	89.1	0.1	(54.7)	24.8	(0.4)	0.2	-	-	-	-	-	-
Second to last hour of test	13:52-14:52	82.3	20	88.4	0.4	(53.8)	24.4	(0.8)	0.4	-	-	-	-	-	-
Period with steam data	15:38-15:56	82.4	7	89.0	0.0	(54.7)	24.8	(0.4)	0.2	10.1	0.2	(620)	281	(16)	7.2

Calculations based on data from the data acquisition system (DAS). Carbon balance used to calculate flue gas flows.

TABLE A-12. VARIATIONS OF PROCESS VARIABLES DURING LOAD FOLLOWING TESTS

Test	Time	Gross Power - MW		FGD Inlet Flow (300°F, 29.92 Hg)				FGD Inlet O_2 Free Flow (300°F, 29.92 Hg)				Corrected steam flow			
		Mean Mean	Std. Jev. Std. dev.	Mean		Std. dev.		Mean		Std. dev.		Mean		Std. dev.	
				(ACFM)	m^3/s	(ACFM)	m^3/s	(ACFM)	m^3/s	(ACFM)	m^3/s	(lb/h)	kg/h	(lb/h)	kg/h
Steady 80 MW load 07/18/79	9:02-11:44	82.69	0.27	(338472)	159.741	(2727)	1.287	(316742)	149.486	(2484)	1.172	(42751)	19392	(1203)	545.7
Steady min. sustainable load (MSL) 7/24/79	12:04-12:56	57.26	0.36	(229727)	108.419	(1632)	0.770	(213763)	100.885	(1578)	0.745	-	-	-	-
Load following 80-57 MW 7/19/79 Load change	10:52-11:32	69.18	9.63	(277319)	130.880	(37649)	17.768	(259056)	122.261	(35360)	16.688	(40994)	18594	(1847)	837.8
After load change	11:34-13:02	60.95	0.16	(254403)	120.065	(2240)	1.057	(237792)	112.225	(2008)	0.948	(39924)	18109	(1543)	699.9
Last hour of test	12:02-13:02	60.99	0.13	(253889)	119.822	(2222)	1.049	(237279)	111.983	(1945)	0.918	(39707)	18011	(1462)	663.2
Load following 57-80 MW 7/24/79	13:02-13:32	68.25	9.54	(286357)	135.145	(39647)	18.711	(266582)	125.813	(37403)	17.652	-	-	-	-
Load change	13:02-13:32	68.25	9.54	(286357)	135.145	(39647)	18.711	(266582)	125.812	(37402)	17.652	-	-	-	-
After load change	13:34-15:56	82.44	0.37	(323645)	152.743	(2973)	1.403	(302933)	142.968	(2852)	1.346	-	-	-	-
Last hour of test	14:56-15:56	82.40	0.21	(324556)	153.173	(1638)	0.773	(303993)	143.469	(1546)	0.730	-	-	-	-
Second to last hour of test	13:52-14:52	82.34	0.33	(323096)	152.484	(1645)	0.776	(302342)	142.690	(1595)	0.753				
Period with steam data	15:38-15:56	82.37	0.23	(324095)	152.956	(1204)	0.568	(303648)	143.306	(1182)	0.558	(51048)	23155	(1331)	603.7

TABLE A-13. VARIATIONS OF PROCESS VARIABLES DURING LOAD FOLLOWING TESTS

Test	Time	Raw Coal Rate				Vol.% O_2 (dry) at air preheater inlet		Vol.% O_2 (dry) at FGD inlet		Vol.% CO_2 (dry) at FGD outlet	
		Mean		Std. Dev.		Mean	Std. Dev.	Mean	Std. Dev.	Mean	Std. Dev.
		(lb/h)	kg/h	(lb/h)	kg/h						
Steady 80 MW load 7/18/79	9:02-11:44	(77381)	35099	(432)	196	3.59	0.08	6.96	0.06	11.81	0.10
Steady min. sustainable load (MSL) 7/24/79	12:04-12:56	(53257)	24157	(295)	134	3.23	0.09	7.51	0.05	12.13	0.05
Load following 80-57 MW 7/19/79 load change	10:52-11:32	(65962)	29920	(9331)	4232	3.42	0.33	7.15	0.20	11.71	0.19
After load change	11:34-13:02	(58092)	26350	(336)	152	3.29	0.14	7.08	0.08	11.80	0.12
Last hour of test	12:02-13:02	(58131)	26368	(311)	141	3.30	0.15	7.10	0.08	11.84	0.12
Load following 57-80 MW 7/24/79 Load change	13:02-13:32	(65119)	29537	(9124)	4138	3.53	0.48	7.48	0.24	11.90	0.36
After load change	13:34-15:56	(75976)	34462	(645)	292	3.32	0.13	6.95	0.09	12.33	0.08
Last hour of test	14:56-15:56	(76355)	34634	(394)	179	3.25	0.06	6.88	0.04	12.36	0.04
Second to last hour of test	13:52-14:52	(75600)	34292	(277)	126	3.35	0.13	6.97	0.08	12.28	0.04
Period with steam data	15:38-15:56	(76291)	34605	(194)	88	3.27	0.06	6.86	0.03	12.37	0.05

TABLE A-14. VARIATIONS OF PROCESS VARIABLES DURING LOAD FOLLOWING TESTS

Test	Time	Inlet SO_2		Inlet SO_2 mass flow rate				Outlet SO_2		Outlet SO_2 mass flow rate			
		Mean	Std. Dev	Mean		Std. Dev.		Mean	Std. Dev.	Mean		Std. Dev.	
		ppmv		(lb/h)	kg/h	(lb/h)	kg/h	ppmv		(lb/h)	kg/h	(lb/h)	kg/h
Steady 80 MW load 7/18/79	9:02-11:44	2156	26	(5058)	2294	(73)	33	216	6	(529)	240	(15)	6.8
Steady min. sustainable load (MSL) 7/24/79	12:04-12:56	2164	18	(3446)	1563	(50)	23	215	3	(364)	165	(7)	3.0
Load following 80-57 MW 7/19/79 load change	10:52-11:32	2218	18	(4267)	1935	(601)	273	241	15	(489)	222	(88)	40.0
After load change	11:34-13:02	2180	21	(3844)	1744	(33)	15	192	6	(356)	161	(13)	5.9
Last hour of test	12:02-13:02	2179	24	(3835)	1740	(32)	14	189	5	(351)	159	(11)	5.0
Load following 57-80 MW 7/24/79 load change	13:02-13:32	2164	53	(4302)	1951	(669)	303	222	7	(467)	212	(67)	30.0
After load change	13:34-15:56	2243	12	(5031)	2282	(55)	25	240	9	(570)	258	(22)	10.0
Last hour of test	14:56-15:56	2250	9	(5061)	2296	(33)	15	231	3	(551)	250	(6)	3.0
Second to last hour of test	13:52-14:52	2237	11	(5008)	2272	(43)	20	245	7	(583)	264	(16)	7.2
Period with steam data	15:38-15:56	2253	12	(5061)	2296	(44)	20	234	2	(556)	252	(6)	3.0

TABLE A-15. FGD PLANT NATURAL GAS CONSUMPTION

Period	Month	Date	Process natural gas 10^6 cf	Process natural gas 10^6 m^3	Incinerator natural gas 10^6 cf	Incinerator natural gas 10^6 m^3	Total natural gas 10^6 cf	Total natural gas 10^6 m^3	Incinerator usage as % of total
14	Oct.	09/29/78-11/03/78	3.505	.09925	0.612	.01733	4.117	.11658	15
	*		3.505	.09925	0.612	.01733	4.117	.11658	15
15	Nov.	11/03/78-12/03/78	6.864	.19437	0.570	.01614	7.433	.21048	8
	*		10.369	.29362	1.182	.03347	11.550	.32706	10
16	Dec.	12/03/78-01/02/79	8.890	.25174	0.557	.01577	4.447	.12592	13
	*		14.259	.40377	1.739	.04924	15.997	.45928	11
17	Jan.	01/02/79-02/01/79	1.443	.04086	0.390	.01104	1.833	.05190	21
	*		15.702	.44463	2.129	.06029	17.830	.50489	12
18	Feb.	02/01/79-03/03/79	0.053	.00150	0.238	.00674	0.290	.00821	80
	*		15.755	.44613	2.367	.06702	18.120	.51310	13
19	March	03/03/79-04/02/79	2.390	.06768	0.406	.01150	2.796	.07917	15
	*		18.145	.51381	2.773	.07852	20.916	.59228	13
20	April	04/02/79-05/02/79	4.315	.12219	0.454	.01286	4.768	.13501	10
	*		22.460	.63560	3.227	.09138	25.684	.72729	13
21	May	05/02/79-06/01/79	4.607	.13046	0.482	.01365	5.088	.14408	9
	*		27.067	.76645	3.709	.10503	30.772	.87137	12
22	June	06/01/79-07/01/79	3.254	.09214	0.475	.01345	3.729	.10559	13
	*		30.321	.85859	4.184	.11848	34.501	.97696	12
23	July	07/01/79-07/31/79	4.735	.13408	0.462	.01308	5.197	.14716	9
	*		35.056	.99268	4.646	.13156	39.698	1.12412	12
24	Aug.	07/31/79-08/30/79	3.870	.10959	0.675	.01911	4.545	.12870	15
	*		38.926	1.1026	5.321	.15067	44.243	1.25282	12
25	Sept.	08/30/79-09/29/79	5.486	.15535	0.771	.02813	6.257	.17718	12
	*		44.412	1.25761	6.092	.17251	50.500	1.43000	12
26	Oct.	09/29/79-10/29/79	4.169	.11805	0.620	.01756	4.789	.13561	13
	*		48.581	1.37566	6.712	.19006	55.289	1.56561	12

*cumulative

TABLE A-16. FGD PLANT NATURAL GAS CONSUMPTION BY MODE

Operating Mode	WEIGHTED AVERAGE			OPERATING PERIOD[1] AVERAGES AND STANDARD DEVIATIONS						
	Process cf/d (m^3/d)	Tail Gas Incinerator cf/d (m^3/d)	Total cf/d (m^3/d)	No.[1] Periods	Process Average cf/d (m^3/d)	Std. Dev. cf/d (m^3/d)	Tail Gas Incinerator		Total	
							Average cf/d (m^3/d)	Std. Dev. cf/d (m^3/d)	Average cf/d (m^3/d)	Std. Dev. cf/d (m^3/d)
1	0	13190 (373.5)	13190 (373.5)	10	0	0	18090 (512.2)	6050 (171)	18090 (512.2)	6050 (171)
2	219810 (6224.3)	17280 (489.3)	237090 (6713.6)	27	216420 (6128.3)	25140 (711.9)	17470 (494.7)	4150 (118)	233890 (6623.0)	27600 (781.5)
3	-	-	-	1	170650 (4832.3)	-	15980 (452.5)	-	186630 (5284.8)	-
4	-	12140 (343.8)	12140 (343.8)	8	0	0	13890 (393.3)	5910 (167)	13890 (393.3)	5910 (167)

[1] Operating periods are those periods of more than 24 hours at one operating mode for which natural gas usages are available.

OPERATING MODE DEFINITIONS

Mode 0 - Boiler down, FGD down
Mode 1 - Boiler up, FGD down
Mode 2 - Boiler up, FGD up, full operation
Mode 3 - Boiler up, FGD up, damper open
Mode 4 - Boiler, FGD up, reduction area down

TABLE A-17. SODA ASH CONSUMPTION*

Month	Dates	Soda Ash Consumed kg(net tons)
Oct. 1978	09/29/78-11/02/78	110,497 (121.8)
Nov.	11/03/78-12/02/78	71,940 (79.3)
Dec.	12/03/78-01/01/79	72,847 (80.3)
Jan. 1979	01/02/79-01/31/79	157,215 (173.3)
Feb.	02/01/79-03/02/79	161,932 (178.5)
March	03/03/79-04/01/79	160,118 (176.5)
April	04/02/79-05/01/79	237,864 (262.2)
May	05/02/79-05/31/79	200,843 (221.4)
June	06/01/79-06/30/79	190,872 (210.4)
July	07/01/79-07/30/79	215,275 (237.3)
August	07/31/79-08/29/79	127,822 (140.9)
Sept.	08/30/79-09/28/79	171,004 (188.5)
Oct. 1979	09/29/79-10/23/79	183,523 (202.3)

*From Allied performance information summaries.

TABLE A-18. HOURLY AVERAGE ELECTRICAL ENERGY CONSUMPTION

Month	Mode 1 kW	Mode 2 kW	Mode 3 kW	Mode 4 kW	All kW	Boiler input Eq. (10^6 Btu), 10^9J
Oct.	715	762	742	0	756	(7.013) 7.394
Nov.	706	741	741	727	-	(7.440) 7.844
Dec.	672	724	703	747	0	(7.087) 7.472
Jan.	603	695	709	0	697	(6.891) 7.266
Feb.						
Mar.	694	695	0	641	694	(6.918) 7.294
Apr.	673	700	714	732	698	(7.000) 7.380
May	690	705	0	713	701	(7.002) 7.382
June	712	746	762	729	730	(7.490) 7.897
July	648	714	0	610	700	(7.054) 7.437
Aug.	561	677	721	654	661	(6.682) 7.045
Sept.	654	692	684	684	683	(6.725) 7.090
Oct.	611	701	717	694	688	(6.868) 7.241
Avg.	673	713	733	694		(7.014) 7.395

TABLE A-19. SULFUR AND SULFATE PRODUCTION*

Month	Dates	Sulfur Recovered (Long Tons)	Sulfur Recovered (Net Tons)	Sulfate Produced (Net Tons)
Oct. 1978	09/29/78-11/02/78	258.7	289.7	42.7
Nov.	11/03/78-12/02/78	510.6	571.9	65.4
Dec.	12/03/78-01/01/79	286.6	321.0	5.1
Jan. 1979	01/02/79-01/31/79	126.3	141.5	0.0
Feb.	02/01/79-03/02/79	8.5	9.5	0.0
March	03/03/79-04/01/79	165.5	185.4	33.4
April	04/02/79-05/01/79	323.3	362.1	6.8
May	05/02/79-05/31/79	332.7	372.6	37.8
June	06/01/79-06/30/79	237.1	265.6	57.4
July	07/01/79-07/30/79	345.3	386.7	59.8
August	07/31/79-08/29/79	251.0	281.1	0.0
Sept.	08/30/79-09/28/79	364.6	408.4	0.0
Oct. 1979	09/29/79-10/23/79	267.4	299.5	0.0

*From Allied performance information summaries.

TABLE A-20. NO_x AS A FUNCTION OF BOILER HEAT INPUT

BHI(1) 10^6 Btu/h	NO_x lb/h	NO_x lb/10^6 Btu	BHI(1) 10^9 J/s	NO_x kg/s	NO_x ng/J
543	160	0.29	573	.0202	120
546	168	.31	576	.0212	130
553	156	.28	583	.0196	120
565	159	.28	596	.0200	120
579	186	.32	610	.0234	140
585	181	.31	617	.0228	130
645	219	.34	680	.0276	150
691	258	.37	728	.0325	160
695	276	.40	733	.0348	170
715	274	.38	754	.0345	160
781	307	.39	823	.0387	170
782	304	.39	824	.0383	170
783	317	.40	825	.0399	170
786	305	.39	829	.0384	170
785	307	.39	828	.0387	170
789	302	.38	832	.0380	160
789	299	.38	832	.0377	160
790	320	.41	833	.0403	180
799	299	.37	842	.0377	160
799	341	.43	842	.0430	180
797	298	.37	840	.0375	160
792	311	.39	835	.0392	170
804	334	.42	848	.0421	180
809	353	.44	853	.0445	190
805	155	.19	849	.0195	82
804	330	.41	848	.0416	180
808	210	.26	852	.0264	110
815	330	.40	859	.0416	170

(continued)

(1) Boiler Heat Input

TABLE A-20 (continued)

BHI 10^6 Btu/h	NO_x lb/h	NO_x lb/10^6 Btu	BHI 10^9 J/s	NO_x kg/s	NO_x ng/J
811	226	.28	855	.0285	120
810	163	.20	854	.0205	86
819	160	.20	864	.0202	86
810	328	.40	854	.0413	170
824	171	.21	869	.0215	90
824	155	.19	869	.0195	82
825	150	.18	870	.0189	77
823	205	.25	868	.0258	110
828	199	.24	873	.0251	100
824	148	.18	869	.0186	77
825	164	.20	870	.0207	86
821	268	.33	866	.0338	140
834	328	.39	879	.0413	170
837	200	.24	882	.0252	100
830	166	.20	875	.0209	86
848	333	.39	894	.0420	170
849	141	.17	895	.0178	73
857	351	.41	904	.0442	180
859	265	.31	906	.0334	130
862	250	.29	909	.0315	120
871	245	.28	918	.0309	120
878	252	.29	926	.0318	120
880	392	.45	928	.0494	190
886	344	.39	934	.0433	170
881	413	.47	929	.0520	200
888	362	.41	936	.0456	180
886	268	.30	934	.0338	130
881	260	.30	929	.0328	130

(continued)

TABLE A-20 (continued)

BHI 10^6 Btu/h	NO_x lb/h	NO_x lb/10^6 Btu	BHI 10^9 J/s	NO_x kg/s	NO_x ng/J
884	295	.33	932	.0372	140
885	300	.34	933	.0378	150
892	404	.45	940	.0509	190
896	397	.44	945	.0500	190
899	381	.42	948	.0480	180
899	335	.37	948	.0422	160
895	351	.39	944	.0442	170
895	332	.37	944	.0418	160
905	391	.43	954	.0493	180
909	361	.40	958	.0455	170
909	345	.38	958	.0435	160
903	295	.33	952	.0372	140
906	563	.62	955	.0709	270
912	405	.44	962	.0510	190
910	353	.39	959	.0445	170
918	384	.42	968	.0484	180
915	350	.38	965	.0441	160
918	507	.55	968	.0639	240
926	492	.53	976	.0620	230
923	428	.46	973	.0539	200
926	339	.37	976	.0427	160
922	291	.32	972	.0367	140
920	531	.58	970	.0669	250
934	353	.38	985	.0445	160
936	387	.41	987	.0488	180
939	380	.40	990	.0479	170
936	374	.40	987	.0471	170

(continued)

TABLE A-20 (continued)

BHI 10^6 Btu/h	NO_x lb/h	NO_x lb/10^6 Btu	BHI 10^9 J/s	NO_x kg/s	NO_x ng/J
949	425	.45	1000	.0535	190
944	404	.43	995	.0509	180
947	417	.44	998	.0525	190
947	320	.34	998	.0403	150
941	365	.39	992	.0460	170
946	340	.36	997	.0428	160
958	425	.44	1010	.0535	190
959	410	.43	1010	.0516	180
958	417	.44	1010	.0525	190
950	355	.37	1000	.0447	160
950	341	.36	1000	.0430	160
953	302	.32	1000	.0380	140
964	434	.45	1020	.0547	190
963	416	.43	1020	.0524	180
965	439	.45	1020	.0553	190
964	418	.43	1020	.0527	180
971	415	.43	1020	.0523	180

Statistics
sample size = n =
mean = $\bar{x}$ =
ρ = std. deviation =
confidence limits = CD =
2ρ =

TABLE A-21. WATER SAMPLE ANALYTICAL RESULTS

Date	Sample	TDS (mg/ℓ)	TSS (mg/ℓ)	SO_4 (mg/ℓ)	Cl (mg/ℓ)	COD (mg/ℓ)	TKN (mg/ℓ)	NO_2-NO_3 (mg/ℓ)
10/21/78	[1]FAP	3194	916	1813	265	125	0.32	1.52
	LW	183	1.3	21.6	9.4	0.8	0.28	0.27
	DW	15	0.0	1.5	0.5	0.0	0.13	0.16
	Net	3011	915	1791	256	123	0.04	1.25
10/28/78	FAP	4316	1160	2823	1050	881	0.00	0.65
	LW	161	0.7	25.5	84	56	0.23	0.11
	Net	4155	1159	2797	966	825	-0.23	0.54
11/03/78	FAP	2956	628	1804	930	767	2.87	0.44
	LW	182	0.7	36	78	51	0.45	0.16
	Net	2774	627	1768	852	716	2.42	0.28
11/09/78	FAP	3240	348	1835	700	506	2.88	0.30
	LW	154	0	26.7	670	56	0.20	0.15
	Net	3086	348	1808	30	450	2.68	0.15
11/14/78	FAP	3446	1008	1498	200	439	0.30	0.11
	LW	179	4	24	6.6	4.7	0.08	0.23
	DW	7.0	0.0	4.1	0.33	0.0	0.20	0.62
	Net	3267	1004	1474	193	434	0.22	-0.12
11/20/78	FAP	3244	272	1750	340	60.8	0.32	1.14
	LW	142	2.7	26.6	42	4.6	0.32	0.31
	DW	0.0	0.0	1.3	0.37	0.0	0.72	0.02
	Net	3102	269	1723	298	56	0.00	0.83
12/02/78	FAP	2638	580	1186	270	304	0.03	0.50
	LW	141	1.3	23.4	6.9	1.6	0.02	0.87
	DW	35	0.0	0.0	0.38	0.0	0.02	0.85
	Net	2497	579	1163	263	302	0.01	-0.37
01/05/79	FAP	3680	524	1674	560	188	0.48	0.46
	DW	157	0.0	0.0	0.38	0	0.42	0.00
01/25/79	FAP	2600	3344	1502	25	302	0.38	0.44
	DW	0.0	0.0	1.0	0.40	1.5	0.30	0.00
03/08/79	FAP	1283	1228	5600	485	276	1.64	1.62
	LW	2225	57	238	10.8	10.8	1.78	0.64
	Net	1058	1171	5362	474	265	-0.14	0.98

(continued)

[1] FAP = fly ash purge
LW = lake water
DW = distilled water
Net = FAP - LW

TABLE A-21 (continued)

Date	Sample	TDS (mg/ℓ)	TSS (mg/ℓ)	SO_4 (mg/ℓ)	Cl (mg/ℓ)	COD (mg/ℓ)	TKN (mg/ℓ)	NO_2-NO_3 (mg/ℓ)
03/26/79	FAP	2054	422	18500	510	59	0.48	0.30
	LW	223	205	215	17.5	7.4	0.34	1.60
	DW	0.05	63.4	40.0	1.15	2.98	0.30	0.07
	Net	1831	217	18280	492	52	0.14	-1.30
04/01/80	FAP	9035	2423	34600	570	3386	61.0	0.00
	LW	7169	6.8	420	13.0	0.0	0.3	0.44
	DW	0.0	0.0	48.0	0.58	0.0	0.24	0.00
	Net	1866	2416	34180	557	3386	60.7	-0.44
04/07/79	FAP	16800	1128	38400	790	8352	2.20	0.00
	LW	186	27	288	12.0	2.8	0.22	0.29
	Net	16610	1101	38110	778	8349	1.98	-0.29
04/13/79	FAP	12513	548	34600	710	15312	4.80	0.00
	LW	311	2.6	326	18	5.6	0.18	0.29
	Net	12200	545	34270	692	15310	4.62	-0.29
04/19/79	FAP	12016	880	515000	660	8256	3.28	0.00
	LW	172	0.0	140	16.5	12.4	0.38	0.00
	Net	11840	880	514900	644	8244	2.90	0.00
05/19/79	FAP	1623	989	41000	568	11248	56	0.00
	LW	161	0.0	190	13.9	82.9	0.4	0.04
	Net	1462	989	40810	554	11160	56	-0.04
06/12/79	FAP	5824	504	18500	445	5376	42	0.00
	LW	242	0.0	210	14.2	23.5	0.8	0.23
	Net	5582	504	18290	431	5352	41	-0.23
06/19/79	FAP	6466	405	40500	220	960	9.0	0.00
	LW	142	28	302	19.0	20.8	3.7	0.42
	Net	6324	377	40200	201	939	5.3	-0.42
06/30/79	FAP	1.89	0.40	10700	279	274	8.0	1.72
	LW	0.13	0.00	211	17.2	11.5	0.7	0.22
	Net	1.76	0.40	10490	262	262	7.3	1.50
07/13/79	FAP	1.57	0.12	7800	450	72	2.10	9.40
	LW	0.14	0.01	200	17.2	5.8	1.75	0.24
	Net	1.43	0.11	7600	433	66	0.35	9.16

(continued)

TABLE A-21 (continued)

Date	Sample	TDS (mg/ℓ)	TSS (mg/ℓ)	SO_4 (mg/ℓ)	Cl (mg/ℓ)	COD (mg/ℓ)	TKN (mg/ℓ)	NO_2-NO_3 (mg/ℓ)
07/18/79	FAP	3.52	0.41	15200	620	432	15.70	0.92
	LW	0.15	0.01	222	16.5	48.7	1.21	0.27
	Net	3.37	0.40	14980	604	383	14.49	0.65
07/21/79	FAP	1.39	0.08	7450	100	344	0.16	0.52
	LW	0.17	0.01	252	11.8	11.1	0.04	0.08
	Net	1.22	0.07	7198	88	333	0.12	0.44
07/22/79	FAP	1.08	0.13	8800	210	468	0.55	0.22
	LW	0.16	0.01	274	13.0	9.6	0.01	0.24
	Net	0.92	0.12	8526	197	458	0.54	-0.02
07/22/79	FAP	1.85	0.26	11200	260	1142	0.41	0.52
	LW	0.20	0.06	252	11.5	1.4	0.03	0.33
	Net	1.65	0.20	10950	248	1141	0.38	0.19
07/25/79	FAP	5.86	1.18	49000	461	620	0.42	0.00
	LW	0.18	0.06	224	13.5	21.2	0.02	0.34
	Net	5.68	1.12	48780	448	599	0.40	-0.34
07/26/79	FAP	5.80	1.36	36000	238	1958	0.25	0.00
	LW	0.20	0.06	245	13.7	8.2	0.05	0.12
	Net	5.60	1.30	35760	224	1950	0.20	-0.12
07/26/79	FAP	2.30	0.35	160000	214	620	0.42	0.43
	LW	0.18	0.01	227	10.4	49	0.04	21
	Net	2.12	0.34	159800	204	571	0.38	-21
09/04/79	FAP	1.65	0.14	13800	52	1714	0.05	0.50
	LW	0.18	0.01	94	13.5	191	0.02	0.73
	Net	1.47	0.13	13710	38	1523	0.03	-0.23
09/22/79	FAP	1.27	2.48	10000	84	724	0.59	0.12
	LW	0.17	.00	254	3.6	159	0.04	0.25
	Net	1.10	2.48	9746	80	565	0.55	-0.12

TABLE A-21a. WATER SAMPLE ANALYTICAL RESULTS - SUMMARY

Statistic and Sample[1]	TDS (mg/ℓ)	TSS (mg/ℓ)	SO_4 (mg/ℓ)	Cl (mg/ℓ)	COD (mg/ℓ)	TKN (mg/ℓ)	NO_2-NO_3 (mg/ℓ)
MEAN[2]							
FAP	3343	597	37674	423	2247	7.5	0.75
LW	369.4	12.5	184	43.5	31.8	0.50	1.11
Net	2989	485	40165	389	2363	7.50	-0.35
DW	26.8	7.9	12.0	0.51	0.56	0.54	0.22
MAX							
FAP	16800	3344	515000	1050	15312	61.0	9.4
LW	7169	205	420	670	191	3.7	21
Net	16610	2416	514900	966	15310	60.7	9.16
DW	157	63.4	48	1.15	2.98	2.20	0.85
MIN.							
FAP	1.08	0.08	1186	25	59	0.0	0.0
LW	0.13	0.0	21.6	3.6	0.0	0.01	0.0
Net	0.92	0.07	1163	30	52	-0.23	-21
DW	0.0	0.0	0.0	0.33	0.0	0.02	0.0
STD. DEV.							
FAP	4320	769	96780	264	3807	16.3	1.74
LW	1362	40.5	111.2	126.6	47.0	0.80	3.99
Net	4275	577	99900	253	3910	16.8	4.52
DW	54.0	22.4	19.9	0.27	1.11	0.70	0.33

[1] FAP = fly ash purge
LW = lake water
DW = distilled water
Net = FAP - LW

[2] Mean based on 29 fly ash purge samples, 27 lake water samples, 27 net values, and 8 distilled water samples.

TABLE A-22. BOILER CAPACITY AND UTILIZATION

	Oct.	Nov.	Dec.	Jan.	Feb.	Mar.	Apr.	May	Jun.	Jul.	Aug.	Sept.	Oct.	All months
Capacity factor[1] FGD down	87	82	77	97	90	82	85	80	62	66	63	71	69	86
Capacity factor FGD full operation	75	77	67	70	-	74	67	75	67	63	70	68	66	70
Utilization parameter, %[2]	95	100	91	100	100	100	100	94	87	110	78	98	82	94

[1] Hours boiler operated/hours in period.

[2] Mwh generated in period/maximum continuous generating capacity (115 Mw/hr).

TABLE A-23. CONTROLLED CONDITIONS TEST SULFUR DATA

Test No.	1	2	3	4	5	6	7	8	9
Generating potential, MW	59.8	60.8	60.7	88.6	88.8	88.3	51.4	49.9	50.3
Gross load, MWe	52.4	53.4	53.3	79.6	79.8	79.3	44.5	43.1	43.5
Sulfur in coal (lb/h) kg/s	(1665) 0.2098	(1787) 0.2252	(1889) 0.2380	(2783) 0.3506	(2278) 0.2870	(2584) 0.3256	(1664) 0.2097	(1664) 0.2097	(1650) 0.2079
Sulfur lost in ash, (lb/h) kg/s	(63) 0.0079	(36) 0.0045	(32) 0.0040	(48) 0.0060	(51) 0.0064	(51) 0.0064	(34) 0.0043	(34) 0.0043	(34) 0.0043
Available sulfur, (lb/h) kg/s	(1602) 0.2018	(1751) 0.2206	(1857) 0.2340	(2738) 0.3450	(2227) 0.2806	(2533) 0.3192	(1630) 0.2054	(1630) 0.2054	(1616) 0.2036
SO_2 out ID fan, ppm wet	2014	2078	2084	2047	2080	-	-	-	-
SO_2 out booster fan, ppm wet	2196	2390	2252	2423	2131	2229	1747	2161	2147
SO_3 out booster fan, ppm wet	0	2.2	1.1	1.2	6.9	1.2	23.6	2.2	36.2
SO_2 out ID fan (lb/h) kg/s	(3147) 0.3965	(3223) 0.4061	(3165) 0.3988	(4755) 0.5989	(4636) 0.5841	- -	- -	- -	- -
SO_2 out booster fan, (lb/h) kg/s	(3397) 0.4280	(3640) 0.4586	(3330) 0.4196	(5544) 0.6985	(4652) 0.5861	(4760) 0.5998	(2572) 0.3241	(3460) 0.14360	(353) 0.0445
SO_3 out booster fan, (lb/h) kg/s	(0) 0	(4) 0.0005	(2) 0.0002	(3) 0.0004	(19) 0.0024	(3) 0.0004	(44) 0.0055	(4) 0.0005	(75) 0.0094
S out ID fans, (lb/h) kg/s	(1575) 0.1984	(1613) 0.2032	(1584) 0.1996	(2380) 0.2999	(2321) 0.2924	- -	- -	- -	- -
S out booster fan, (lb/h) kg/s	(1701) .2143	(1824) .2298	(1668) .2102	(2777) .3499	(2337) .2944	(2385) .3005	(1305) .1644	(1734) .2185	(1798) .2265
S balance closure, S out ID fan/available S	98	92	85	87	104	-	-	-	-
SO_2 out absorber, analyzer, ppm wet	220	216	197	226	213	-	-	-	-
SO_2 out absorber, Method 8, ppm wet	171	184	169	172	192	134	384	174	147
% SO_2 removal, analyzer	88.5	89.0	90.0	88.3	89.3	-	-	-	-
% SO_2 removal, Method 8	91.7	91.7	91.9	92.3	90.4	93.6	76.0	91.2	92.7
SO_3 out absorber, ppm wet	2.0	0.9	2.3	1.0	1.8	1.9	0	0	1.9

TABLE A-24. CONTROLLED CONDITION TEST FLUE GAS CHARACTERISTICS

Test No.	1	2	3	4	5	6	7	8	9
Generating potential, MW	59.8	60.8	60.7	88.6	88.8	88.3	51.4	49.9	50.3
Gross load, MWe	52.4	53.4	53.3	79.6	79.8	79.3	44.5	43.1	43.5
Temperature out FD fans (^{o}F) ^{o}C	(318) 158.9	(339) 170.5	(336) 168.9	(332) 166.6	(323) 161.6	(321) 160.5	- -	- -	- -
Temperature and booster fan, (^{o}F) ^{o}C	(320) 160.0	(336) 168.9	(338) 170.0	(332) 166.6	(294) 145.5	(284) 140.0	(308) 153.3	(278) 136.6	(280) 137.8
Temperature out absorber, (^{o}F) ^{o}C	(125) 51.7	(127) 52.8	(127) 52.8	(127) 52.8	(125) 51.7	(126) 52.2	(128) 53.3	(128) 53.3	(127) 52.8
Static pressure out ID fans ("W.C.) KPa	(2.3) 0.6	(2.7) 0.7	(2.6) 0.6	(3.4) 0.8	(3.6) 0.9	(3.6) 0.9	- -	- -	- -
Static pressure booster fan, ("W.C.) KPa	(15) 3.7	(16) 4.0	(15) 3.7	(16) 4.0	(16) 4.0	(16) 4.0	- -	- -	- -
Flue gas flow in, measured (acfm) m^3/s	(310,600) 146.6	(222,400) 105.0	(216,200) 102.0	(333,700) 157.5	(238,700) 112.6	(333,300) 157.3	(224,500) 106.0	(235,000) 110.9	(255,100) 120.4
Flue gas flow in, calculated (acfm) m^3s	(225,400) 106.4	(223,800) 105.6	(219,000) 103.4	(335,100) 158.1	(321,600) 151.8	(313,200) 147.8	(215,900) 101.9	(232,900) 109.9	(236,300) 111.5
Flue gas flow out, measured (acfm) m^3/s	(233,500) 110.2	(238,300) 112.5	(175,500) 82.83	(331,000) 156.2	(324,400) 153.1	(279,700) 132.0	(191,700) 90.47	(214,400) 101.2	(232,700) 109.8
Flue gas flow out, calculated (acfm) m^3/s	(236,900) 111.8	(237,000) 111.8	(231,200) 109.1	(356,400) 168.2	(336,500) 158.8	(329,700) 155.6	(232,400) 109.7	(252,000) 118.9	(254,800) 120.2
O_2 inlet air preheater, vol.% dry stationary gases out ID fan	3.2	3.2	3.5	3.5	3.4	3.4	-	-	-
$-O_2$, vol.% dry	7.2	9.3	7.5	6.1	6.6	7.1	8.8	8.3	10.0
$-CO_2$ vol.% dry	12.9	12.1	12.8	12.7	12.9	13.1	10.8	10.0	10.8
$-H_2O$, vol.% wet	7.8	6.7	6.2	6.6	6.8	7.0	6.2	6.4	7.5
Stationary gases out absorber									
$-O_2$, vol.% dry	7.2	8.7	7.5	6.7	6.6	6.9	8.9	8.4	9.6
$-CO_2$ vol.% dry	12.9	12.9	12.8	12.1	12.9	13.3	11.4	10.6	10.3
$-H_2O$, vol.% dry, measured	10.2	11.7	11.2	13.5	10.2	10.5	10.2	10.8	10.1
$-H_2O$ vol.% dry, calculated	13.3	13.7	13.7	13.7	12.9	13.3	14.4	14.4	14.0

TABLE A-25. CONTROLLED CONDITION TEST COAL QUALITY

Test No.	1	2	3	4	5	6	7	8	9	Mean	Mean Standard Deviation	95% Confidence limits of mean
Gross load, MWe												
High heating value, (Btu/lb) 10^3 J/kg	(10735) 24952	(10853) 25227	(10813) 25134	(11265) 26184	(11003) 25575	(10968) 25494		(10925) 25394		(10940) 25429	(170) 395	(10780-11100) 25057-25801
Moisture, wt.%	13.72	12.87	12.75	10.88	13.41	13.01		12.14		12.68	0.94	10.19-15.17
Volatile matter, wt.%	34.32	35.27	35.66	37.00	35.32	35.27		34.84		35.38	0.83	34.51-36.23
Fixed carbon, wt.%	42.40	41.73	40.73	42.37	42.48	42.26		42.57		42.08	0.65	41.48-42.68
Ash, wt.%	9.56	10.13	10.86	9.75	8.79	9.46		10.45		9.86	0.69	8.94-10.78
Carbon, wt.%	59.92	60.39	59.57	62.30	60.85	60.62		60.53		60.60	0.87	60.04-61.16
Hydrogen, wt.%	4.15	4.16	4.20	4.40	4.37	4.26		4.20		4.25	0.10	4.16-4.34
Oxygen, wt.%	8.52	8.08	8.14	8.05	8.88	8.48		7.90		8.29	0.35	7.97-8.61
Nitrogen, wt.%	1.18	1.16	1.01	0.95	0.82	0.93		1.30		1.05	0.17	0.89-1.21
Chlorine, wt.%	0.03	0.04	0.06	0.07	0.05	0.04		0.05		0.05	0.01	0.04-0.06
Sulfur, wt.%	2.92	3.17	3.41	3.60	2.83	3.20		3.43		3.22	0.28	2.96-3.48
Coal rate (lb/h) kg/s	(57020) 7.184	(56360) 7.101	(55400) 6.980	(77300) 9.740	(80480) 10.14	(80760) 10.18	(48500) 6.111	(48500) 6.111	(48100) 6.060	- -	- -	- -

TABLE A-26. COMPARISON OF JULY 1979 BOILER DRAFT LOSSES WITH BASELINE I OR II

	46 Baseline I	44 July 1979 Flow Tests	53 July 1979 Flow Tests	92 Baseline I	92 Baseline II	80 July 1979 Flow Tests
NIPSCO Gross MW	45.4	43.7	53.0	90.0	92.6	79.6
NIPSCO raw coal rate (lb/h)	(43700)	(48370)	(56260)	(81000)	(83300)	(79510)
kg/s	5.506	6.095	7.089	10.20	10.50	10.02
Air, Air Heater (In W.C.)	(2.0)	0.0	0.0	4.3	2.3	1.9
Pa	500			1100	570	470
Ducts and dampers, A.P.H. to burners, burners and windbox (In W.C.)	(1.0)	0.0	0.0	2.3	1.1	0.2
Pa	250			570	270	50
Total air resistance (In W.C.)	(3.0)	0.0	0.0	6.6	3.4	2.1
Pa	750			1600	850	520
Furnace and convection banks (In W.C.)	(2.4)	1.0	0.8	3.1	2.8	2.4
Pa	600	250	200	770	700	600
Flues to A.H. (In W.C.)	(0.6)	0.0	0.1	1.0	0.9	0.5
Pa	150		25	250	220	120
Gas, Air heater (In W.C.)	(3.0)	2.5	2.5	4.6	2.9	3.0
Pa	750	620	620	1100	720	750
Total gas resistance (In W.C.)	(6.0)	3.5	3.4	8.7	6.6	5.9
Pa	1500	870	850	2200	1600	1500
Total boiler resistance (In W.C.)	(9.0)	3.5	3.4	15.3	10.0	8.0
Pa	2200	870	850	3810	2490	2000

TABLE A-27. BOILER DRAFT LOSSES DURING JULY 1979 FLOW TESTS

Test Date	1 07/21/79	2 07/22/79	3 07/22/79	Avg. 1-3	4 07/25/79	5 07/26/79	6 07/26/79	Avg. 4-6	7 07/28/79	8 07/28/79	9 07/28/79	Avg. 7-9
NIPSCO Gross MW	52.36	53.40	53.33	53.03	79.60	79.80	79.33	79.58	44.50	43.14	45.50	43.71
NIPSCO raw coal rate (lb/h)	(57020)	(56360)	(55400)	(56260)	(77300)	(80480)	(80760)	(79513)	(48500)	(48500)	(48100)	(48367)
kg/s	7.184	7.101	6.980	7.089	9.740	10.14	10.18	10.018	6.111	6.111	6.060	6.094
Air, Air heater (In. W.C.)	(0.0)	0.0	0.0	0.0	2.0	2.0	1.8	1.9	0.0	0.0	0.0	0.0
Pa	0.0	0.0	0.0	0.0	500	500	450	470	0.0	0.0	0.0	0.0
Ducts and dampers, A.P.H. to burners, Burners and windbox (In. W.C.)	(0.0)	0.0	0.0	0.0	0.0	0.0	0.5	0.2	0.0	0.0	0.0	0.0
Pa	0.0	0.0	0.0	0.0	0.0	0.0	120	50	0.0	0.0	0.0	0.0
Total air resistance (In. W.C.)	(0.0)	0.0	0.0	0.0	2.0	2.0	2.3	2.1	0.0	0.0	0.0	0.0
Pa	0.0	0.0	0.0	0.0	500	500	570	520	0.0	0.0	0.0	0.0
Furnace and convection banks, (In. W.C.)	(0.6)	1.0	0.9	0.8	2.4	2.4	2.3	2.4	1.0	1.0	1.0	1.0
Pa	150	250	220	200	600	600	570	600	250	250	250	250
Flues to A. H. (In. W.C.)	(0.3)	0.0	0.0	0.1	0.5	0.5	0.6	0.5	0.0	0.0	0.0	0.0
Pa	75	0.0	0.0	25	120	120	150	120	0.0	0.0	0.0	0.0
Gas, air heater (In. W.C.)	(2.5)	2.5	2.5	2.5	3.0	3.0	3.1	3.0	2.5	2.5	2.5	2.5
Pa	620	620	620	620	750	750	770	750	620	620	620	620
Total gas resistance (In. W.C.)	(3.4)	3.5	3.4	3.4	5.9	5.9	6.0	5.9	3.5	3.5	3.5	3.5
Pa	850	870	850	850	1500	1500	1500	1500	870	870	870	870
Total boiler resistance (In. W. C.)	(3.4)	3.5	3.4	3.4	7.9	7.9	8.3	8.0	3.5	3.5	3.5	3.5
Pa	850	870	850	850	2000	2000	2100	2000	870	870	870	870

APPENDIX B

BOILER DESCRIPTION

APPENDIX B

BOILER DESCRIPTION

LOCATION

The boiler, Mitchell No. 11, is one of four pulverized coal fired boilers at the D. H. Mitchell Station located near the extreme northwest city limits of Gary, Indiana and fronts on Lake Michigan near its extreme southern end. Mitchell No. 11 is the easternmost boiler at the station. Lake Michigan is the source and receiving body for the cooling water. Elevation at grade is 586.5 feet above mean sea level. Elevation above normal lake level at grade is eight feet. Atmospheric temperatures are as follows:

Summer average 23.4°C (75°F), Peak 37.8°C (100°F)

Winter average -1.1°C (30°F), Low -13.3°C (-8°F)

The FGD plant accepts the total flue gas from the discharge of two ID fans. The orientation of the boiler and its flue gas duct system relative to the FGD plant is shown on Figure B-1. The original ductwork at the discharge of the ID fans ran westward to a stack shared with Mitchell No. 6. For retrofit of the FGD plant part of this ductwork was replaced and reoriented 180° to the suction of the FGD plant booster fan. A louvered bypass damper was installed in this section of duct. The original duct to the breeching of the existing stack was kept in place.

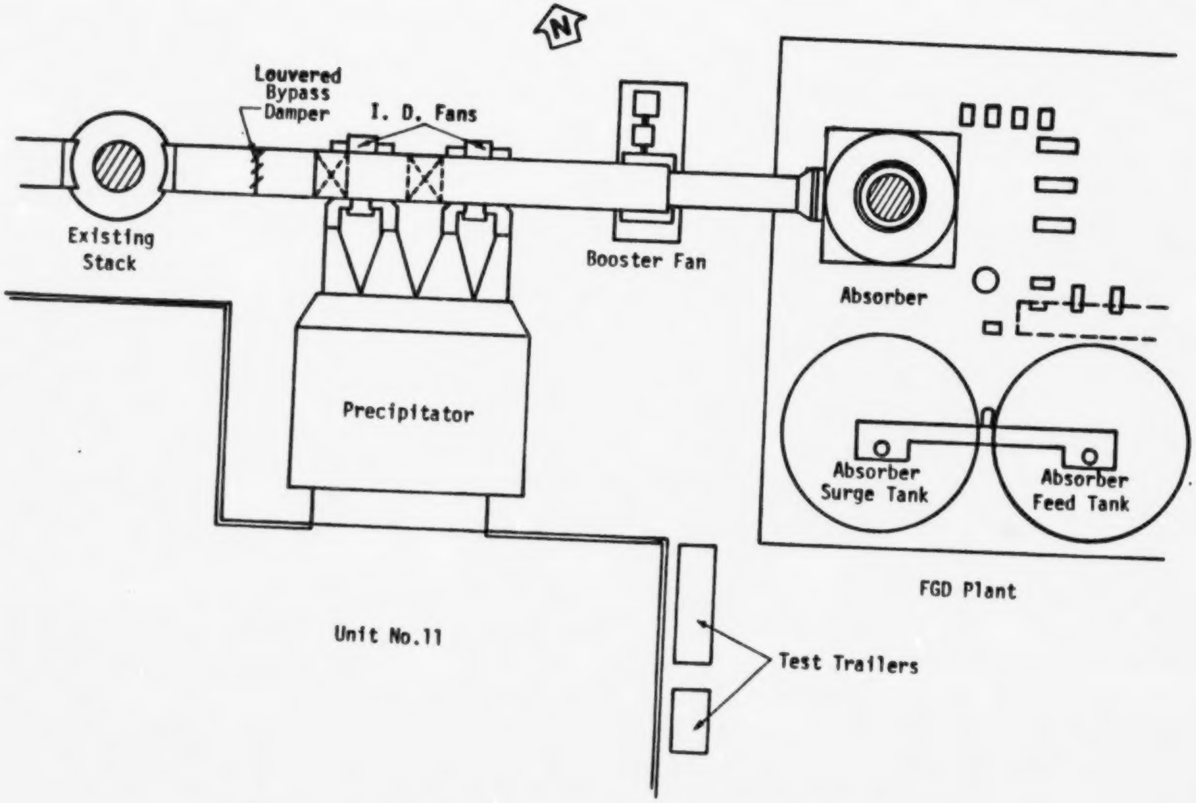
Louvered
Bypass
Damper
I. D. Fans
N
Existing
Stack
Booster Fan
Absorber
Precipitator
Absorber
Surge Tank
Absorber
Feed Tank
FGD Plant
Unit No.11
Test Trailers

Figure B-1. Plan view of boiler and FGD plant.

TYPE AND EXPECTED CONDITIONS

Mitchell No. 11 is a balanced draft wet bottom radiant wall-fired reheat boiler which was manufactured by Babcock and Wilcox and placed into service in 1970. The unit can develop a maximum continuous steam capacity of 103 kg/s (821,000 lbs/h) with a corresponding turbo-generator output of 115.6 megawatts. The steam parameters controlled are 12,893 kPa (1870 psi), 540.6°C, (1005°F) secondary superheater outlet, 540.6°C (1005°F) reheat outlet. A flow configuration of the boiler unit and major auxiliaries is shown on Figure B-2.

Fuel

Coal from the Captain mine was burned during the latter part of the first demonstration year and during the second demonstration year. It was obtained from the No. 6 bed located in Southwest Illinois. It is a high volatile bituminous, H.V.C. rank (ASTM D388 designation). Sulfur content was within the inlet sulfur design range of the FGD plant. Based on the good coal preparation and firing performance of this coal, it appears to be quite compatible with Mitchell No. 11 design.

FUEL HANDLING

Coal is transferred from rail cars or storage piles through a crusher to screens which classify to one inch maximum. This coal is transported by conveyor to a bunker of about 1,814,000 kg (2000 tons) capacity. Coal flows by gravity through four fill pipes, 24 inch diameter, to four coal feeder scales. Raw coal was sampled from the gravity feed pipes. The four fill pipes run full and thus act as a

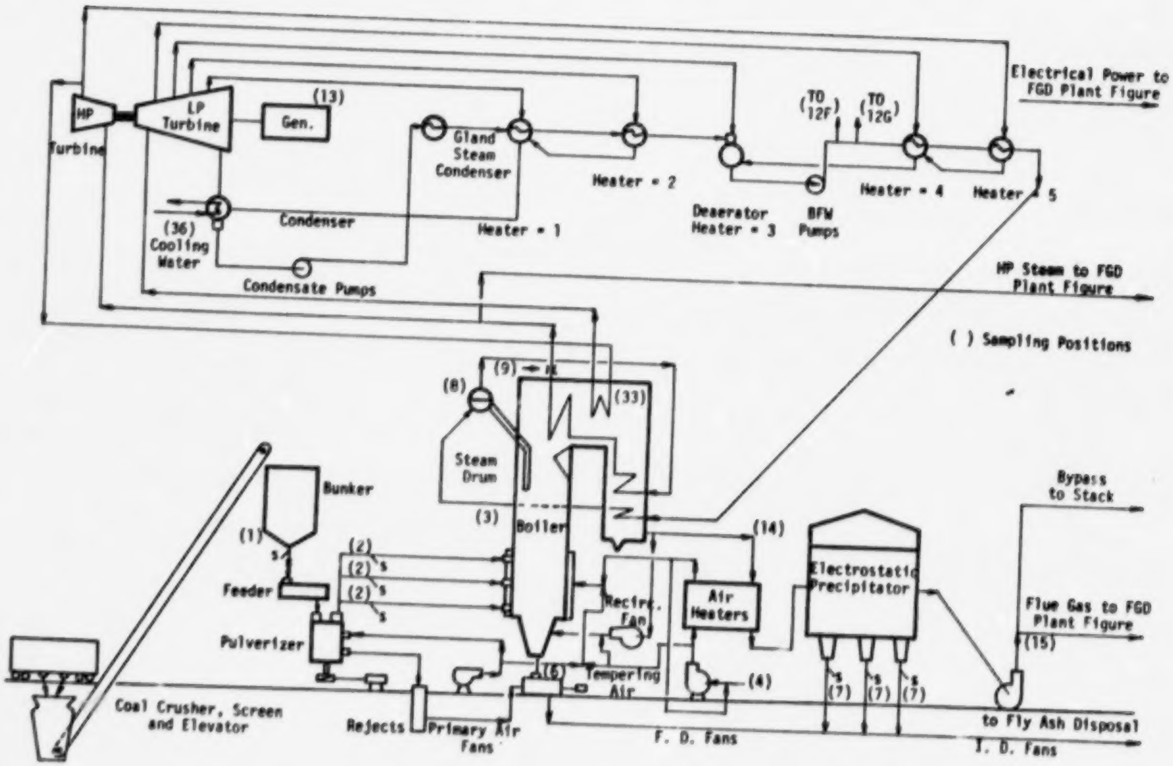
HP
Turbine
LP
Turbine
(13)
Gen.
Gland
Steam
Condenser
Heater # 1
Heater # 2
Deaerator
Heater # 3
BFW
Pumps
TO
12F
TO
12G
Heater # 4
Heater # 5
Electrical Power to
FGD Plant Figure
(36)
Cooling
Water
Condenser
Condensate Pumps
HP Steam to FGD
Plant Figure
( ) Sampling Positions
(8)
(9)
(33)
Steam
Drum
(3)
Boiler
Bunker
(1)
s
Feeder
(2)
(2)
(2)
s
Pulverizer
Coal Crusher, Screen
and Elevator
Rejects
Primary Air
Fans
(6)
Recirc.
Fan
Tempering
Air
(14)
Air
Heaters
(4)
F. D. Fans
Electrostatic
Precipitator
(7)
(7)
(7)
Bypass
to Stack
Flue Gas to FGD
Plant Figure
(15)
to Fly Ash Disposal
I. D. Fans

Figure B-2. Mitchell No. 11 boiler.

seal for pressurization of the feeders and pulverizers. The raw coal flows across four weigh feeders and then by gravity to four pulverizers.

FUEL FIRING

General

The boiler front wall contains 12 circular burners. Each of four pulverizers feed coal-air (coal, 77% through 200 U. S. mesh) to three burners. Excess combustion air is normally 19% (design value) but higher at low loads. Each coal pulverizer receives coal from a separate gravimetric coal feeder mounted on the operating floor between the overhead coal bunkers and the coal pulverizers. At low load, the option exists of turning off one or more feeders.

The normal operating capacity of Mitchell No. 11 ranges from 46 to 115 megawatts. During normal operating conditions, the load is changed at the rate of 1.0 megawatt per minute. The load can be changed from a hot initial condition at the faster rate of 3.3 megawatts per minute up or down if required. From a warm initial condition, the load can be changed at the rate of 1.6 megawatts per minute up or down if required.

Equipment Descriptions

Coal Feeders (4)--

These units are Stock Equipment gravimetric coal feeders, pressurized, which gravimetrically measure coal feed rate. Maximum capacity of each is 7.6 kg/s (60,000 lbs/h).

Pulverizers (4)--

B&W Model EL70 ball mills, ball and race type, medium speed. New ball diameters are 11 1/4" and wear is to 7". Primary air is supplied by a blower for each mill. Power input, including blower, is 14 kWh/ton coal.

Burners (12)--

B&W circular register, 36" diameter with steaming capacities of 11.5 kg/s (91,200 lb/h), each burner. Igniters are gas fired. The fuel feeding pattern is shown on Figure B-3. Connections for sampling "as fired" coal were used during the test program.

Air Supply--

Two forced draft fans are provided as well as four primary fuel-air fans. See Table B-1 for fan information. In case of draft unbalance due to malfunction, the boiler trips at one inch H_2O positive pressure with a two second time delay.

TABLE B-1. FANS AND PUMPS INFORMATION

Forced Draft Fans:	
Manufacturer	American Standard
Number of Fans	2
Capacity at 132°F	203,000 c.f.m.
Discharge Static Head	17.3 inches H_2O
Driver	A.C. 700 H.P. - 1185 R.P.M.

TABLE B-1 (concluded)

Induced Draft Fans:	
Manufacturer	American Standard
Number of Fans	2
Capacity at 289°F	228,000 c.f.m.
Static Head	-18.2 inches H_2O
Driver	Elliot 900 H.P. - 887 R.P.M.
Variable Speed Drive	American Blower Corporation
Boiler Feed Pumps:	
Manufacturer	Ingersoll-Rand Company
Number	2
Type	10-Stage Barrel
Capacity	452,000 lb/h (1013 gpm)
Net Developed Head	2096 psi
Variable Speed Drive	American Blower Corporation
Driver	West. 1750 H.P. - 3573 R.P.M.

COMBUSTION CONTROL

The 721 analog system supplied by Bailey Meter is used.

Fuel Control

1. MW output change affects turbine demand.
2. Turbine demand change affects steam control valves.

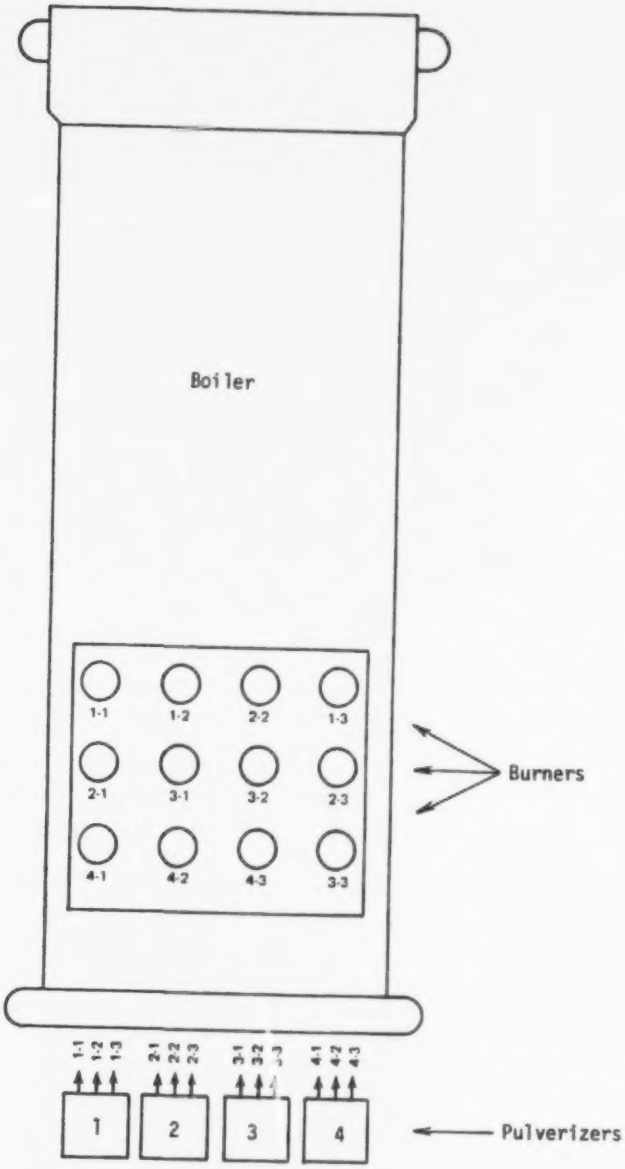
Boiler
1-1
1-2
2-2
1-3
2-1
3-1
3-2
2-3
4-1
4-2
4-3
3-3
Burners
1-1
1-2
1-3
2-1
2-2
2-3
3-1
3-2
3-3
4-1
4-2
4-3
1
2
3
4
Pulverizers

Figure B-3. Fuel distribution to burners.

3. Change in steam control valve affects boiler pressure (measured as secondary superheater outlet pressure) and steam flow rates.
4. Change in summation of steam header pressure and flow rate affects mill demand.

Air Control

Primary combustion air flow is controlled by summation of steam flow and header pressure. Relevant sensors include air flow, air temperature, flue gas oxygen, steam temperature, steam flow and header pressure. Secondary combustion air is manually controlled by the control room operator. Figure B-4 is typical of the "load ramp" used for controlling excess air.

FIRE SIDE CHARACTERISTICS

The furnace configuration is shown in the elevation drawing (Figure B-5). The furnace volume is 76,126 cu. ft. with effective heating surface as follows:

Reheat surface - (15,818 sq. ft.) 1469.5 m^2
Superheater - (54,130 sq. ft.) 5028 m^2
Reheater - (18,226 sq. ft.) 1693.3 m^2
Economizer - (9,550 sq. ft.) 887 m^2
Air heater - (132,200 sq. ft.) 12281 m^2*

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*Decreased for second demonstration year.

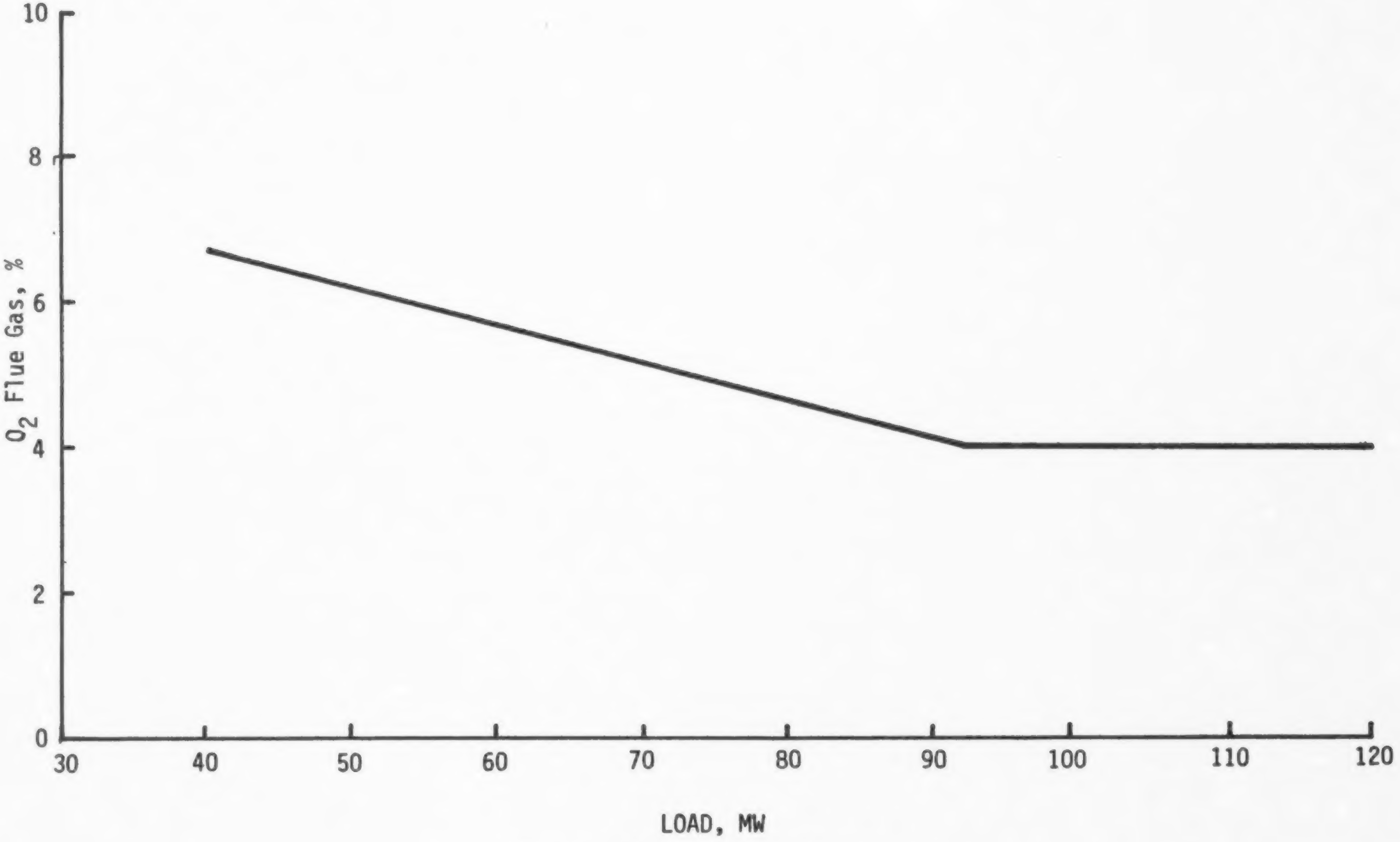

Figure B-4. Boiler load ramp.

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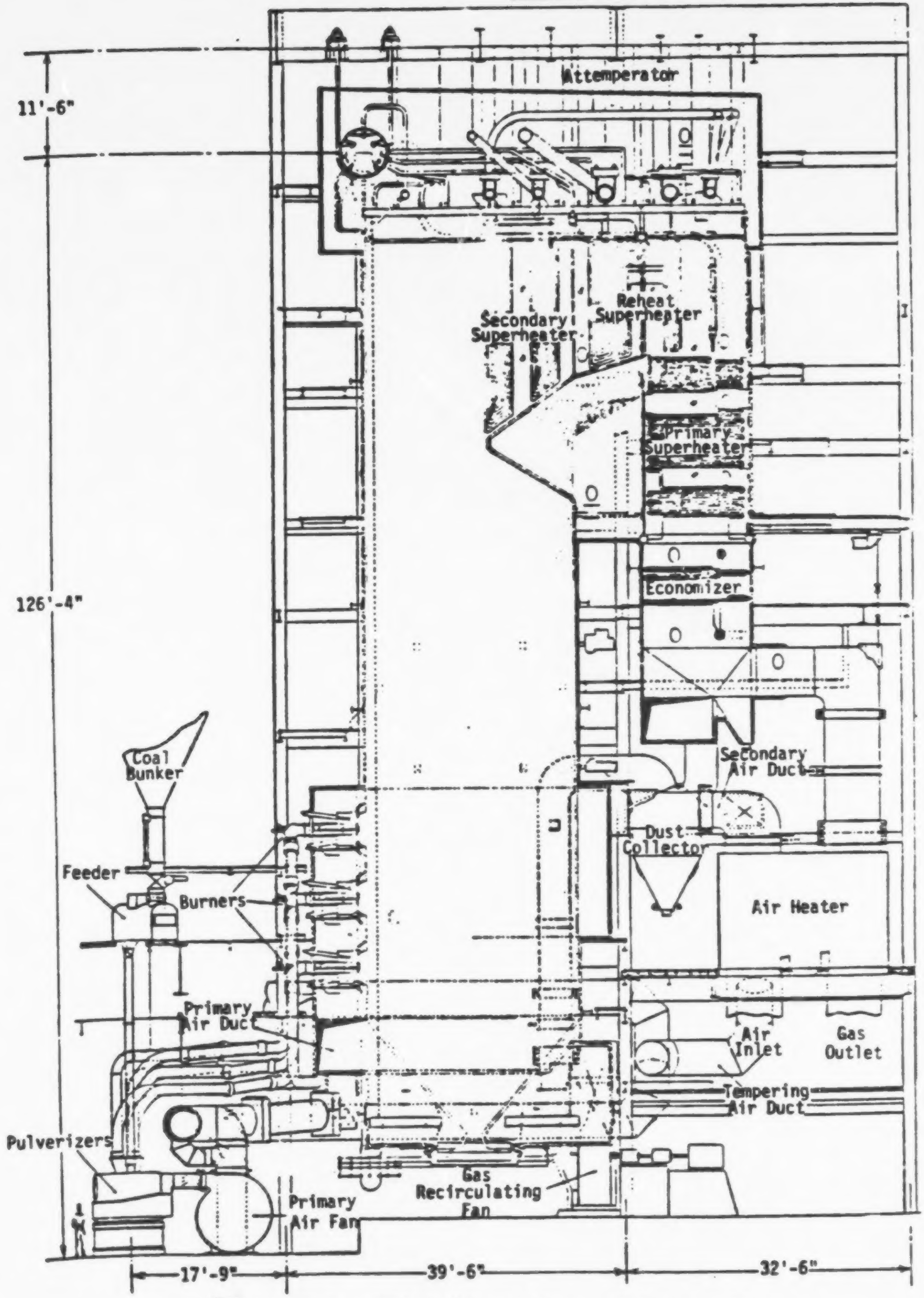

Attemperator
11'-6"
Secondary Superheater
Reheat Superheater
Primary Superheater
Economizer
126'-4"
Coal Bunker
Feeder
Burners
Secondary Air Duct
Dust Collector
Air Heater
Primary Air Duct
Air Inlet
Gas Outlet
Tempering Air Duct
Pulverizers
Primary Air Fan
Gas Recirculating Fan
17'-9"
39'-6"
32'-6"

Figure B-5. Mitchell station - Unit no. 11.

Nominal flue gas temperatures at full load are:

Furnace exit - 1248.9°C (2280°F)

Superheater exit - 1037.8°C (1900°F)

Entering reheater - 987.8°C (1810°F)

Leaving reheater - 698.9°C (1290°F)

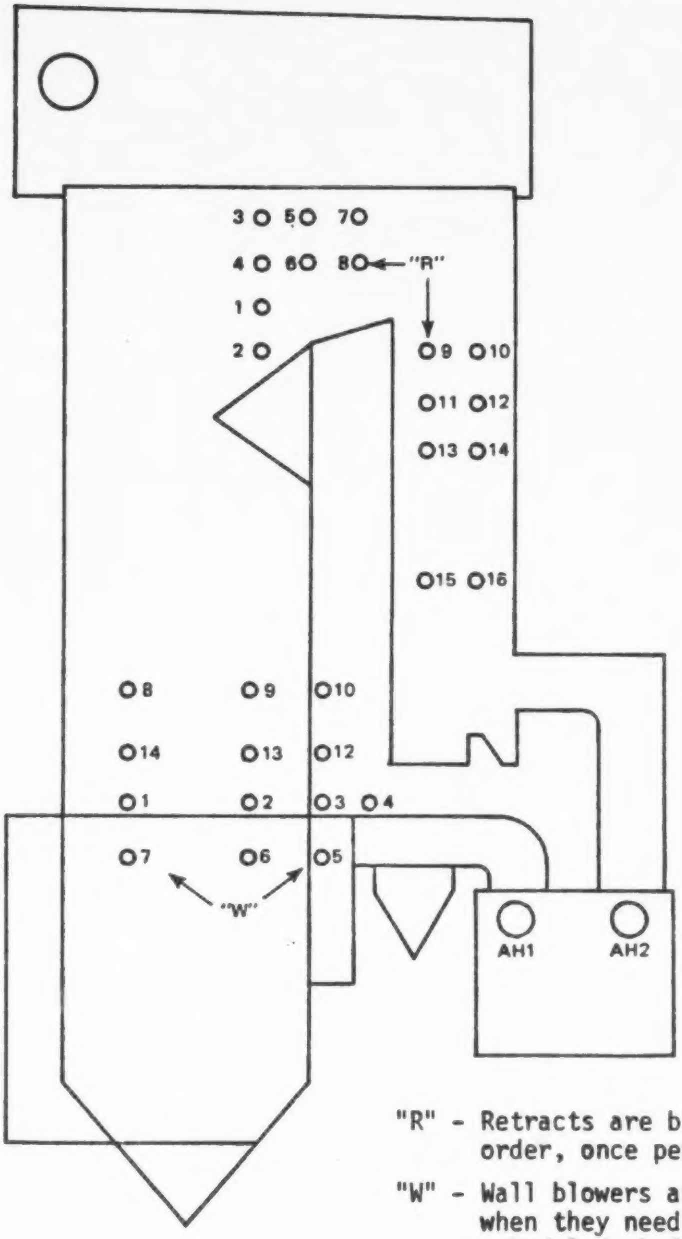
Leaving economizer - 368.3°C (695°F)

Gas recirculation from outlet the economizer, used mainly for startups, is provided.

The boiler is provided with a United Conveyor Co. ash handling system with a water-flooded ash hopper beneath the boiler. Pyrites rejection is combined with bottom ash and sluiced to the ash pond. Dust collector hoppers are emptied through rotary feeders and pneumatically conveyed out of the unit, and sluiced to the ash pond. Means for obtaining representative samples of pyrites and bottom ash were not available. The boiler is provided with a Copes-Vulcan Div. air energized wall deslagger and retractable soot blower system operating on a programmed daily cycle. The cycle is initiated by the control room operator. Wall blowers may be actuated independently at any time, to maintain steam superheat/reheat temperatures. Figure B-6 shows the relative positions of retracts and wall blowers. The normal soot blowing cycle requires about two hours to execute and is performed once a day

STEAM SIDE CHARACTERISTICS

Steam conditions, superheat and reheat vs. load have been summarized below. Steam for FGD consumption was withdrawn from the system at the secondary superheater outlet, after the main steam flowmeter. Nominal steam conditions were:


3
5
7
4
6
8
"R"
1
2
9
10
11
12
13
14
15
16
8
9
10
14
13
12
1
2
3
4
7
6
5
"W"
AH1
AH2

"R" - Retracts are blown in order, once per shift.

"W" - Wall blowers are blown when they need it or if scheduled, before retracts, in order.

Even numbered are right; odd numbered are left.

Figure B-6. Soot blowers.

Pressure	12411 kPa (1800 psig)
Temperature	537.8°C (1000°F)
Enthalpy	3443.4 kJ/kg (1480.4 Btu/lb)

Heating surface areas are 5028 m^2 (54,130 sq. ft.) (superheater), 1693.3 m^2 (18,226 sq. ft.) (reheater). Boiler feed water pumps have been described in Table B-1. Condensate is heated by a series of two water heat exchangers, deaerated, heating continued in two extraction-steam heaters, and then fed to the economizer. The secondary superheater outlet steam rate is as follows:

Full load	2.1 kg/kWh (6.8 lb/kWh)
92 MWG	2.97 kg/kWh (6.55 lb/kWh)
46 MWG	2.88 kg/kWh (6.35 lb/kWh)

Attemperation spray water is withdrawn after the BFW rate meter and enters (1) hot reheat steam and (2) primary superheated steam as controlled by the relevant sensors and controls. Nominal condenser exhaust pressure is 5.1 kPa (1.5" Hg). The condenser is a horizontal single-pass Ingersoll Rand unit with vertically divided water boxes, of phosphorized admiralty construction. Tube length is 30 ft. and surface area is 5852.9 m^2 (63,000 sq. ft.). Once through cooling water from Lake Michigan is used.

The turbine-generator set is made by General Electric Co., is of the tandem compound single-flow reheat type, carries a nameplate rating of 115.6 MW, runs at 3600 RPM and has 5 extraction openings.

Air Preheaters

Two Ljungstrom type notary air heaters in parallel are provided. The design condition shown on Table B-2 do not apply during the second demonstration year because some of the heat transfer baskets were removed to provide higher feed temperatures to the booster blower. Air inleakage to the flue gas is typical of this type of air heater.

TABLE B-2. REGENERATIVE AIR HEATER (DESIGN CONDITIONS)

MAKE: AIR PREHEATER CORP. TYPE: LJUNGSTROM SIZE: 22-VIR-66
HEATING SURFACE/UNIT SQ. FT.: 66,100 NO./BOILER: 2

% of Full Load	100	53
Evaporation, Lb. Steam/h	821,000	437,400
Ambient Air Temp. °F	80	80
Air Leaving Heater Lb/h	900,000	491,000
Air Entering Heater Lb/h	991,500	562,000
Gas Entering Heater Lb/h	1,091,000	633,000
Gas Leaving Heater Lb/h	1,182,500	704,000
Temp. of Air Entering °F	100	100
Temp. of Air Leaving °F.	602	607
Temp. of Gas Leaving Uncorrected °F.	303	294
Temp. of Gas Leaving Corrected °F.	289	276
Resistance, Air Side in. H_2O	3.55	1.25
Resistance, Gas Side in. H_2O	5.80	2.35
Total Pressure Diff. Across Heater in. H_2O	16.85	9.10

ELECTROSTATIC PRECIPITATOR

A cold end electrostatic precipitator, manufactured by American Standard Co., is in the flue gas stream between the air heaters and the induced draft fans.

Design Changes

The information that has been supplied was obtained from boiler design information. We are not aware of any significant design changes except as follows:

- Coal feed chutes were enlarged during the first demonstration year to minimize coal feed problems.
- Prior to the second demonstration year some of the air preheater baskets that provide heat storage were removed to raise the temperature at inlet of the FGD plant.

Early in 1977, a serious accident occurred that required extensive repair to the boiler, the flue gas ducts and the air preheater. The repairs were made without substantial design change. However, there are indications the damage might have resulted in increased air inleakage in the air preheater.

APPENDIX C

FGD PLANT DESCRIPTION

APPENDIX C

FGD PLANT DESCRIPTION

The processes are proprietary designs of Davy McKee (formerly Wellman-Powergas, Inc.) and Allied Chemical Corporation. The integrated process is referred to in this report as the Wellmann-Lord/Allied Chemical FGD process or WL/Allied FGD process. Logical separation of the various process steps are:

SO_2 absorption - Davy McKee

SO_2 recovery and scrubber solution
regeneration - Davy McKee

Purge treatment - Davy McKee

SO_2 reduction - Allied Chemical

The description to follow is based on Davy McKee and Allied Chemical non-proprietary design data obtained primarily from the Davy McKee Project Manual prepared during the design phase (Phase I) of the project.[13]

The WL/Allied FGD process removes SO_2 from the flue gas stream by scrubbing with an aqueous sodium sulfite/bisulfite solution and subsequent thermal regeneration to recover the SO_2. The liberated SO_2 is then reduced to elemental sulfur which is sold. The FGD unit was designed to remove 90% of the SO_2 delivered with the flue gas at flue gas rates equivalent to a boiler load of 92 MW (80% of full boiler load). The absorber is designed to take up to about 183 m^3/s (388,000 acfm) of flue gas, equivalent to 110 MW of load, but this rate can be sustained for only a limited time because of limited capacity of the solution regeneration part of the FGD plant.

The critical design criteria are approximately as follows:

Flue gas temperature,	148.9°C (300°F)
Flue gas pressure,	101 kPa (14.7 psia)
Maximum flue gas flow,	183 m^3/s (388,000 acfm)
Gross MW equivalent,	110 MW
Steam equivalent,	94.4 kg/s (749,000 lb/h)
Design flue gas flow,	151 m^3/s (320,000 acfm)
Gross MW equivalent,	92 MW
Steam equivalent,	76.0 kg/s (603,000 lb/h)
Inlet SO_2 at design flow,	0.6101 kg/s (4,842 lb/h)
Equivalent SO_2 concentration,	2,185 ppmv

Any combination of flue gas volume and inlet SO_2 concentration that results in an SO_2 feed rate greater than about 0.63 kg/s (5,000 lbs/h) for periods up to 83 hours is excess capacity for the recovery area. This means that sustained operation at excess capacity would lower the performance level to below 90% SO_2 recovery. The absorber and the recovery area have the turndown capability for steady state operation down to 46 MW boiler load. However, the lower limit for sustained operation of the reduction area is higher than 46 MW due to operating characteristics of the reduction system.

The diagram (Figure C-1) shows the process steps. The FGD plant accepts the total flue gas stream from the discharge of the boiler's

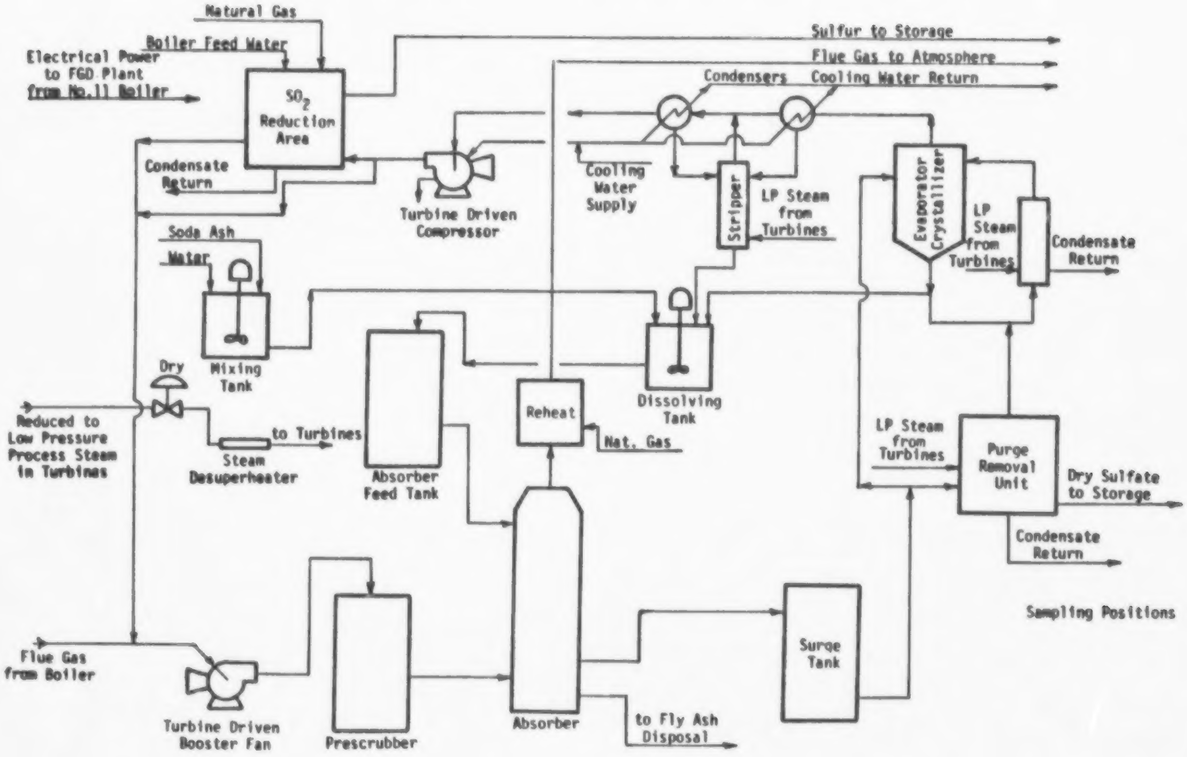

Natural Gas
Boiler Feed Water
Electrical Power to FGD Plant from No.11 Boiler
SO_2 Reduction Area
Condensate Return
Turbine Driven Compressor
Sulfur to Storage
Flue Gas to Atmosphere
Condensers
Cooling Water Return
Cooling Water Supply
Stripper
LP Steam from Turbines
Evaporator Crystallizer
LP Steam from Turbines
Condensate Return
Soda Ash
Water
Mixing Tank
Dry
Reduced to Low Pressure Process Steam in Turbines
Steam Desuperheater
to Turbines
Absorber Feed Tank
Reheat
Nat. Gas
Dissolving Tank
LP Steam from Turbines
Purge Removal Unit
Dry Sulfate to Storage
Condensate Return
Sampling Positions
Flue Gas from Boiler
Turbine Driven Booster Fan
Prescrubber
Absorber
to Fly Ash Disposal
Surge Tank

Figure C-1. Schematic diagram of FGD plant.

induced draft (ID) fans using a steam turbine driven booster blower to force the flue gas stream through the prescrubber and absorber. The flue gas is admixed with recycle tail gas which is exhausted from the SO_2 reduction unit.

The prescrubber is a single-stage orifice contactor for removing additional particulate matter. A pump recirculates the scrubber water from a sump back to the contactor. In order to control a solids buildup in the liquid stream, a purge stream is withdrawn; makeup water is added to the prescrubber to compensate for this loss and to humidify the flue gas. This purge stream is sent to the power station's fly ash settling ponds.

The cooled, humidified flue gas leaves the prescrubber and enters the bottom of a three-stage absorber.

A demister is provided at the base of the absorber to prevent the flue gases from carrying undue amounts of entrained solution from the prescrubber.

The absorber consists of multiple contact stages for sulfur dioxide absorption. An absorber demister pad is provided above the top contact stage to prevent entrained absorber solution from being exhausted with the treated gases. The sulfur dioxide-rich gases pass upwardly through the contact stages and top demister pad and are exhausted from the system. Each contact stage of the absorber has a separate recirculation system to promote good gas-liquid contact. The absorbing process is conducted at about 54.4°C (130°F), and spent absorbing solution is withdrawn from the bottom contact stage. Lean absorbing solution is added to the top contact stage. The process chemistry is as follows:

$$SO_2 + Na_2SO_3 + H_2O \longrightarrow 2NaHSO_3$$

The bisulfite-rich solution is then passed to the evaporator-crystallizer unit. A slipstream of spent absorber solution is passed through the purge treatment system to remove the Na_2SO_4 for eventual disposal.

The evaporator is a single effect forced circulation unit. The heat exchanger employs steam and the clean condensate is discharged for reuse in the boiler system. The heat supplied to the liquor decomposes the sodium bisulfite solution to the sulfite (which crystalizes out), sulfur dioxide, and water:

$$2\,NaHSO_3 \xrightarrow{heat} Na_2SO_3 + SO_2 + H_2O$$

The sodium sulfite slurry which is drawn from the evaporator circulating loop, and the stripped condensate are combined in the agitated dissolving tank. This solution from the dissolving tank is recycled to the top stage of the absorber as the lean absorbing solution after the addition of makeup sodium carbonate to compensate for the sodium values lost in the purge.

The products of desorption, sulfur dioxide and water vapor, are exhausted from the evaporator and passed through the primary condenser. The condensate from the primary condenser is sent to the condensate stripper and non-condensed vapors are passed through the secondary condenser. Also passed to the secondary condenser is the overhead vapor from the condensate stripper. The condensate from the secondary condenser is also returned to the condensate stripper. The outlet gas from the secondary condenser is removed from the recovery system by the steam turbine driven compressor, and the sulfur dioxide is passed to the SO_2 reduction unit. Tanks for the bisulfite solution drawn from the absorber and for the sulfite feed solution provide a limited amount of surge capacity for operating at flue gas rates in excess of design.

The purge treatment unit is designed to minimize the amount of purge and to produce a salable byproduct. Purge requirements are reduced by subjecting part of the absorber product stream to a fractional crystalization process. Sodium sulfate rich crystals are produced by chilling the solution under very specific conditions. This treatment allows the majority of the active sodium compounds (sulfite and bisulfite) to remain in the mother liquor which is separated from the crystals by a centrifuge. This mother liquor is then returned to the main process and the crystals are dehydrated in a drying system. The final product is a dry solid which is shipped from the plant for use as makeup in the Kraft pulping industry.

The SO_2 reduction is accomplished in a two-stage process employing a primary catalytic reaction of SO_2 with natural gas to produce sulfur and some hydrogen sulfide followed by a secondary Claus conversion system in which the hydrogen sulfide is reacted with residual SO_2 to produce additional sulfur. The primary reaction system consists of two packed-bed regenerative heat exchangers and a catalyst-packed reduction reactor. These vessels and their connecting flues are refractory-lined for protection against high temperatures and corrosive gases.

Essentially half of the SO_2 in the feed is reduced to elemental sulfur by direct reaction with the reducing gas:

$$2SO_2 + CH_4 \longrightarrow CO_2 + 2H_2O + 2S$$

Simultenaously, a major portion of the remaining SO_2 is converted into hydrogen sulfide (H_2S) and additional sulfur by a similar reaction:

$$3SO_2 + 2CH_4 \longrightarrow 2CO_2 + 2H_2O + 2H_2S + S$$

The gas flows and operating conditions in the primary reaction system are carefully controlled so that the mixture of H_2S and unconverted SO_2 in the product gases from the primary reaction system closely approximates the ideal volumetric ratio (two parts H_2S to one part SO_2) required for the subsequent Claus reaction:

$$2H_2S + SO_2 \xrightarrow{\qquad\qquad} 2H_2O + 3S$$

At the same time, maximum utilization of the reducing agent is achieved and the formation of undesirable side-reaction products is minimized.

The function of the regenerative heat exchangers is to remove the heat from the gases leaving the reactor, and to utilize this heat to raise the temperature of the gases entering the reactor. At appropriate intervals, the duties of the regenerative heat exchangers are alternated by reversing the flows through these two vessels. This is accomplished by changing the position of the four-way valve which is operated in conjunction with four butterfly valves.

During the first half of the cycle, the mixture of concentrated SO_2 gas from the Wellman-Lord SO_2 Recovery Process and methane enters the bottom of a regenerative heat exchanger. The packing in this unit is hot from the previous half cycle, so the gas passing through is heated to the temperature required for the primary reactions. During this half cycle, two butterfly valves are open and two are closed. The gas flows downward through the reactor where the reactions occur, then down through the second regenerative heat exchanger where the gas gives up heat to the packing.

When the temperatures in the regenerative heaters reach predetermined levels, the four-way valve and the four butterfly valves are repositioned to reverse the flows through these vessels.

Since the sulfur dewpoint of the gas leaving the primary reaction system is quite high, it is extremely important that the relatively cold SO_2 gas stream from the Wellman-Lord SO_2 Recovery Process not create any colder areas in the four-way valve or the lower sections of the regenerative heat exchangers where sulfur may condense when the flow direction is reversed. Consequently, the incoming SO_2-CH_4 mixture is passed through a heat exchanger where part of the heat in the hot gases leaving the primary reactor system is used to raise the temperature of the incoming gases to a level where premature sulfur condensation is avoided.

Since the hot gases from the primary reaction system would be highly corrosive at high temperatures to metals used in the heat exchanger, a portion of these gases is first passed through a sulfur condenser and then remixed with the remaining portion so that the sulfur content and the temperature of the mixture are lower. The tempered gas mixture is used to preheat the SO_2-methane-feed in the heat exchanger. Elemental sulfur that condenses in the sulfur condenser flows by gravity to the sulfur storage. The sulfur-bearing gases leaving the heat exchanger are subsequently passed through a second condenser before being sent to the Claus conversion system. Additional elemental sulfur is removed in this condenser and transferred to the sulfur storage.

The Claus conversion system is a two-stage unit. Associated with the Claus unit is an interstage condenser from which elemental sulfur also passes to storage. A final condenser follows the second stage converter to recover the last portion of the elemental sulfur product.

The tail gases from the SO_2 reduction unit pass through an incinerator where air is introduced to oxidize the remaining H_2S to SO_2.

Natural gas is also burned in the incinerator to raise the temperature and accelerate the H_2S oxidation.

The hot gases from the incinerator are admixed with the untreated flue gases at inlet of the booster blower.

Operating procedures cannot be discussed because of the proprietary nature of the processes. An SO_2 analyzer was used to monitor the SO_2 removal performance and make corrections. Process controls were incorporated without consideration for testing needs. Some needed measurements such as scrubber solution and fly ash purge flows were difficult to determine.

APPENDIX D

CARBON BALANCE FLUE GAS FLOW CALCULATION

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CARBON BALANCE FLUE GAS FLOW CALCULATION

This section describes a method for calculating the flue gas flow from the CO_2 concentration of the flue gas. However, if CO_2 data are not available, an alternative is to calculate the flue gas flow from the O_2 concentration. Either the O_2 concentration or the CO_2 concentration may be used as the independent variable because, on a dry flue gas basis, setting the O_2 concentration also sets the CO_2 concentration (Figure E-1).

The method which will be described uses the CO_2 concentration at the outlet of the FGD unit to calculate the scrubber outlet flow first, followed by the inlet flow calculation. Inlet flow was generally more useful for performance analysis.

The flue gas flow calculation has been programmed for a computer. The data input requirements, output variables, and steps in the flue gas calculation are as follows:

Stream Identification

X - Coal
Y - Air
ZI - Scrubber Inlet
ZO - Scrubber Outlet

Carbon Balance Input Variables

Coal Rate and Coal Analysis

WCX = ash-free coal feed rate to boiler, lb/h

PCX = % carbon in coal by weight on an ash-free basis

PSX = % sulfur in coal by weight on an ash-free basis

PWX = % moisture in coal by weight on an ash-free basis

PHX = % hydrogen in coal by weight on an ash-free basis

POX = % oxygen in coal by weight on an ash-free basis

PNX = % nitrogen in coal by weight on an ash-free basis

Flue Gas Analysis

PCO2ZO = % CO_2 by volume (dry basis) in the flue gas downstream from the scrubber

Temperatures

TY = ambient air temperature, ^oF

TZI = scrubber inlet temperature, ^oF

TZO = scrubber outlet temperature, ^oF

Process

PSAT = percent saturation of combustion and leakage air

PCL = percent by weight of carbon in coal that is not oxidized (about 0.5%)

SAE = sulfur absorption efficiency (about 90%)

Carbon Balance Output Variables

QZI = wet flue gas flow rate to scrubber, ACFM (101.3 kPa (29.92" Hg), TZI)

QZO = wet flue gas flow rate leaving scrubber, ACFM (101.3 kPA (29.92" Hg), TZO)

QZI300 = wet flue gas flow rate to scrubber, ACFM (101.3 kPa (29.92" Hg), 148.9°C (300°F))

QZI130 = wet flue gas flow rate leaving scrubber, ACFM (101.3 kPa (29.92" Hg), 54.4°C (130°F)

PH2OZI = the percent water by volume in the flue gas entering the scrubber

PH2OZO = the percent water by volume in the flue gas leaving the scrubber

PO2Z1 = the percent oxygen by volume in the flue gas entering the scrubber on a dry basis

PO2ZO = the percent oxygen by volume in the flue gas leaving the scrubber on a dry basis

PCO2ZI = the percent carbon dioxide by volume in the flue gas entering the scrubber on a dry basis

(Note: The percent carbon dioxide by volume in the flue gas leaving the scrubber, PCO2ZO, is an input variable).

EXO2 = The percent excess air.

* = multiplication sign

I. Carbon Balance

$\underline{CO_2 \text{ from Tailgas}}$

$$WSO2 = \frac{PSX}{100}\ \frac{\#S}{\#\ Coal} * WCX\ \frac{\#\ Coal}{h} * \frac{64.07\ \#\ SO_2}{32.07\ \#\ S} = \frac{PSX*WCX*64.07}{100.*32.07}\ \frac{\#SO_2}{h}$$

the moles of SO_2 from the coal.

$$WSO2A = WSO2\ \frac{\#\ SO_2}{h} * \frac{SAE}{100}\ \frac{\#\ SO_2}{\#\ SO_2}\ \frac{\text{absorbed}}{\text{from coal}}$$

$$= \frac{WSO2 * SAE}{100}\ \frac{\#\ SO_2 \text{absorbed}}{h},$$ the moles SO_2 absorbed by the FGD unit.

$$MCO2TG = WSO2A\ \frac{\#\ SO_2\text{absorbed}}{h} * .008267\ \frac{\#\ \text{moles}\ CO_2}{\#\ SO_2}$$

$$= .008267 * WSO2A\ \frac{\#\ \text{moles}\ CO_2}{h},$$ the moles of CO_2 from the tail gas.

$\underline{CO_2 \text{ from Soda Ash}}$

$$MCO2SA = 530\ \frac{\#\ Na_2CO_3}{h} * \frac{\#\ \text{mole}\ Na_2CO_3}{105.99\ \#\ Na_2CO_3} * \frac{1\ \#\ \text{mole}\ CO_2}{\#\ \text{mole}\ Na_2CO_3}$$

$$= \frac{530}{105.99}\ \frac{\#\ \text{moles}\ CO_2}{h},$$ the moles of CO_2 from the soda ash.

$\underline{CO_2 \text{ from Coal}}$

$$MCO2X = MCO2X1\ \frac{\#\ \text{moles C}}{h} *\left(1 - \frac{PCL}{100}\right) \frac{\#\ \text{moles}\ CO_2}{\#\ \text{mole C}}$$

$$= MCO2X1 *\left(1 - \frac{PCL}{100}\right) \frac{\#\ \text{moles}\ CO_2}{h},$$ the moles of CO_2 from the coal.

$\underline{\text{Total } CO_2}$

$$MCO2ZO = MCO2TG + MCO2SA + MCO2X\ \frac{\#\ \text{moles}\ CO_2}{h},$$ the total moles of CO_2 leaving the scrubber.

II. Calculate Moles of Dry Flue Gas Leaving Scrubber

$$MDRYZO = MCO2ZO\ \frac{\#\ \text{moles}\ CO_2}{h} * \frac{100}{PCO2ZO}\ \frac{\#\ \text{moles dry flue gas}}{\#\ \text{moles}\ CO_2}$$

$$= \frac{MCO2ZO}{PCO2ZO} * 100\ \frac{\#\ \text{moles dry gas}}{h},$$ the molar flow of dry flue gas

leaving the scrubber.

Set T = TZO, and Calculate the Molar Humidity of the Flue Gas

$A1 = .0383002$
$A2 = .00316537 * T$
$A3 = 1.12837E\text{-}4 * (T)^2$
$A4 = 2.01743E\text{-}6 * (T)^3$
$A5 = 2.03101E\text{-}8 * (T)^4$
$A6 = 1.05016E\text{-}10 * (T)^5$
$A7 = 2.29667E\text{-}13 * (T)^6$

$$\text{SATHUM} = A1-A2+A3-A4+A5-A6+A7 \ \frac{\#\ H_2O}{\#\ \text{dry air}}$$

$$\text{MHUM} = \text{SATHUM} \ \frac{\#\ H_2O}{\#\ \text{dry air}} * \frac{28.98\ \#\ \text{air}}{\#\ \text{mole air}} * \frac{\#\ \text{mole}\ H_2O}{18.016\ \#H_2O}$$

$$= \text{SATHUM} * \frac{28.97}{18.016} \ \frac{\#\ \text{moles}\ H_2O}{\#\ \text{mole dry air}}, \text{ the molar humidity.}$$

Note: This polynomial is accurate to within ±3% over the range 50-180°F.

IV. Calculate Moles of Wet Flue Gas Leaving Scrubber

$$\text{MH2OZO} = \text{MDRYZO} \ \frac{\#\ \text{moles dry air}}{h} * \text{MHUM} \ \frac{\#\ \text{moles}\ H_2O}{\#\ \text{mole dry air}}$$

$$= \text{MDRYZO} * \text{MHUM} \ \frac{\#\ \text{moles}\ H_2O}{h}, \text{ the moles of water leaving the scrubber.}$$

$$\text{MTOTZO} = \text{MDRYZO} + \text{MH2OZO} \ \frac{\#\ \text{moles}}{h}, \text{ the total molar}$$

flow of wet gas leaving the scrubber.

V. Calculate Wet Flue Gas Flow from Scrubber

$$\text{MVO} = 359.05 \ \frac{ft^3}{\#\ \text{mole}} * \frac{(\text{TZO} + 460)\ ^{\circ}R}{492\ ^{\circ}R}, \text{ the gas molar volume at the}$$

scrubber exit temperature.

$$\text{QZO} = \text{MTOTZO} \ \frac{\#\ \text{moles}}{h} * \text{MVO} \ \frac{ft^3}{\#\ \text{mole}} * \frac{1\ h}{60\ \text{min}}$$

$$= \frac{\text{MTOTZO} * \text{MVO}}{60} \ \frac{ft^3}{\text{min}} \text{ at TZO, the total wet gas flow leaving the scrubber.}$$

$$\text{MV130} = 359.05 \ \frac{ft^3}{\#\ \text{mole}} * \frac{(130 + 460)\ ^{\circ}R}{492^{\circ}R}, \text{ the gas molar volume at } 54.4^{\circ}C\ (130^{\circ}F)$$

$$\text{QZO130} = \text{MTOTZO} \ \frac{\#\ \text{moles}}{h} * \text{MV130} \ \frac{ft^3}{\#\ \text{mole}} * \frac{1\ h}{60\ \text{min}}$$

$= \frac{\text{MTOTZO * MV130}}{60} \frac{\text{ft}^3}{\text{min}}$, the total flow of wet flue gas leaving the scrubber at 54.4°C (130°F)

$\text{PH2OZO} = \frac{\text{MH2OZO} \frac{\text{\# moles } H_2O}{\text{h}}}{\text{MTOTZO} \frac{\text{\# moles}}{\text{h}}} * 100 = \frac{\text{100 * MH2OZO}}{\text{MTOTZO}}$, the percent water in the scrubber outlet gas by volume.

VI. Oxygen Balance

$\text{MSO2X} = \text{WSO2} \frac{\text{\# } SO_2}{\text{h}} * \frac{\text{\# mole } SO_2}{\text{64.07 \# } SO_2} = \frac{\text{WSO2}}{64.07} \frac{\text{\# moles } SO_2}{\text{h}}$, the moles of SO_2 from the coal.

$\text{MH2OX} = \frac{\text{PWX}}{100} \frac{\text{\# } H_2O}{\text{\# coal}} * \text{WCX} \frac{\text{\# coal}}{\text{h}} * \frac{\text{\# mole } H_2O}{\text{18.016 \# } H_2O}$

$= \frac{\text{PWX * WCX}}{\text{100 * 18.016}} \frac{\text{\# moles } H_2O}{\text{h}}$, the moles of water from the coal.

$\text{MHX} = \frac{\text{PHX}}{100} \frac{\text{\# hydrogen}}{\text{\# coal}} * \text{WCX} \frac{\text{\# coal}}{\text{h}} * \frac{\text{\# mole H}}{\text{1.008 \# H}}$

$= \frac{\text{PHX * WCX}}{\text{100 * 1.008}} \frac{\text{\# moles H}}{\text{h}}$, the moles of hydrogen (as H) from the coal.

$\text{MO2X} = \frac{\text{POX}}{100} \frac{\text{\# oxygen}}{\text{\# coal}} * \text{WCX} \frac{\text{\# coal}}{\text{h}} * \frac{\text{\# mole } O_2}{\text{32.0 \# } O_2}$

$= \frac{\text{POX * WCX}}{\text{100 * 32.00}} \frac{\text{\# moles } O_2}{\text{hr}}$, the moles of oxygen from the coal.

$\text{MN2X} = \frac{\text{PNX}}{100} \frac{\text{\# nitrogen}}{\text{\# coal}} * \text{WCX} \frac{\text{\# coal}}{\text{h}} * \frac{\text{\# mole } N_2}{\text{28.08 \# } N_2}$

$= \frac{\text{PNX * WCX}}{\text{100 * 28.02}} \frac{\text{\# moles } N_2}{\text{h}}$, the moles of nitrogen from the coal.

$\text{MO2RQ} = \text{MCO2X} \frac{\text{\# moles } CO_2}{\text{h}} + \text{MSO2X} \frac{\text{\# moles } SO_2}{\text{h}}$

$+ \text{MHX} \frac{\text{\# moles H}}{\text{h}} * \frac{\text{1 \# mole } O_2}{\text{4 \# moles H}} - \text{MO2X} \frac{\text{\# moles } O_2}{\text{h}}$,

the moles of oxygen required for combustion of the coal.

$\text{MAIRRQ} = \text{MO2RQ} \frac{\text{\# moles } O_2}{\text{h}} * \frac{\text{\# mole air}}{\text{.21 \# mole } O_2} = \frac{\text{MO2RQ}}{.21} \frac{\text{\# moles air}}{\text{h}}$,

the moles of air entering with the combustion oxygen.

$\text{MN2RQ} = \text{MAIRRQ} \frac{\text{\# moles air}}{\text{h}} * \frac{\text{.79 \# moles } N_2}{\text{\# mole air}}$

$$= .79 * \text{MAIRRQ} \ \frac{\#\text{ moles } N_2}{h}$$, the moles of nitrogen entering with the combustion air.

VII. Calculate Excess Air in Flue Gas

$$\text{MSO2ZO} = \frac{\left[\text{WSO2} \ \frac{\#\ SO_2}{h} - \text{WSO2A} \ \frac{\#\ SO_2\text{absorbed}}{h}\right]}{64.07 \ \frac{\#\ SO_2}{\#\text{ mole } SO_2}}$$

$$= \frac{[\text{WSO2} - \text{WSO2A}]}{64.07} \ \frac{\#\text{ moles } SO_2}{h}$$, the moles of SO_2 leaving the FGD unit.

$$\text{MAIRZO} = \text{MDRYZO} \ \frac{\#\text{ moles dry flue gas}}{h} - \left[\text{MCO2ZO} \ \frac{\#\text{ moles } CO_2}{h} + \text{MSO2ZO} \ \frac{\#\text{ moles } SO_2}{h} + \text{MN2X} \ \frac{\#\text{ moles } N_2}{h} + \text{MN2RQ} \ \frac{\#\text{ moles } N_2}{h}\right]$$, the moles of dry air leaving the FGD unit.

$$\text{MO27O} = \text{MAIRZO} \ \frac{\#\text{ moles air}}{h} * \frac{.21\ \#\text{ mole } O_2}{\#\text{ mole air}}$$

$$= .21 * \text{MAIRZO} \ \frac{\#\text{ moles } O_2}{h}$$, the moles of oxygen leaving the FGD unit.

$$\text{PO2ZO} = \frac{\text{MO2ZO} \ \frac{\#\text{ moles } O_2}{h}}{\text{MDRYZO} \ \frac{\#\text{ moles dry gas}}{h}} * 100$$

$$= \frac{\text{MO2ZO} * 100}{\text{MDRYZO}}$$, the percent oxygen by volume in the scrubber outlet gas on a dry basis.

VIII. Calculate Moles Dry Flue Gas Entering Scrubber

$$\text{MSO2A} = \frac{\text{WSO2A} \ \frac{\#\ SO_2\text{absorbed}}{h}}{64.07 \ \frac{\#\ SO_2}{\#\text{ mole } SO_2}} = \frac{\text{WSO2A}}{64.07} \ \frac{\#\text{ moles } SO_2\text{absorbed}}{h}$$, the moles of SO_2 absorbed by the scrubber.

$$\text{MDRYZI} = \text{MDRYZO} \ \frac{\#\text{ moles dry gas}}{h} + \text{MSO2A} \ \frac{\#\text{ moles } SO_2\text{ absorbed}}{h} - (\text{MCO2TG} + \text{MCO2SA}) \ \frac{\#\text{ moles } CO_2}{}$$, the molar flow of dry gas entering the scrubber

IX. Calculate Moles of Combustion Air

$$MAIRY = MAIRRQ\ \frac{\#\ moles\ air}{h} + MAIRZO\ \frac{\#\ moles\ air}{h},$$ the molar flow of combustion and leakage air.

X. Set T = TY, and Calculate the Molar Humidity of the Ambient Air from the polynomial in Step III.

XI. Water Balance

$$MH2OY = MAIRY\ \frac{\#\ moles\ air}{h} * MHUM\ \frac{\#\ moles\ H_2O}{\#\ mole\ dry\ air} * \frac{PSAT}{100}$$

$$= \frac{MAIRY * MHUM * PSAT}{100}\ \frac{\#\ moles\ H_2O}{h},$$ the moles of water in the combustion and leakage air.

$$MH2OZI = MH2OY\ \frac{\#\ moles\ H_2O}{h} + MH2OX\ \frac{\#\ moles\ H_2O}{h}$$

$$+ MHX\ \frac{\#\ moles\ H}{h} * \frac{1\ \#\ mole\ H_2O}{2\ \#\ moles\ H},$$

the total moles of water entering the scrubber.

XII. Calculate Moles Wet Flue Gas Entering Scrubber

$$MTOTZI = MH2OZI\ \frac{\#\ moles\ H_2O}{h} + MDRYZI\ \frac{\#\ moles\ dry\ gas}{h},$$

the total molar flow of wet gas entering the scrubber.

XIII. Calculate Wet Flue Gas Flow Entering Scrubber

$$MVI = 359.05\ \frac{ft^3}{\#\ mole} * \frac{(TZI + 460)\ ^{o}R}{492\ ^{o}R},$$ the gas molar volume at the scrubber inlet temperature.

$$QZI = MTOTZI\ \frac{\#\ moles}{h} * MVI\ \frac{ft^3}{\#\ mole} * \frac{1\ h}{60\ min}$$

$$= \frac{MTOTZI * MVI}{60}\ \frac{ft^3}{min}$$ at TZI, the total wet gas flow entering the scrubber.

$$MV300 = 359.05\ \frac{ft^3}{\#\ mole} * \frac{(300 + 460)\ ^{o}R}{492^{o}R},$$ the gas molar volume at 148.9°C (300°F).

$$QZI300 = MTOTZI\ \frac{\#\ moles}{h} * MV300\ \frac{ft^3}{\#\ mole} * \frac{1\ h}{60\ min}$$

$$= \frac{\text{MTOTZI * MV300}}{60}\ \frac{ft^3}{min}$$, the flue gas flow at 148.9°C (300°F).

XIV. Calculate Inlet Flue Gas Composition

$$\text{PCO2ZI} = \frac{\text{MCO2ZO}\ \frac{\#\ \text{moles}\ CO_2}{h} * 100}{\text{MDRYZI}\ \frac{\#\ \text{moles dry gas}}{h}}$$

$$= \frac{\text{MCO2ZO * 100}}{\text{MDRYZI}}$$, the present carbon dioxide by volume in the scrubber inlet gas on a dry basis.

$$\text{PO2ZI} = \frac{\text{MO2ZO}\ \frac{\#\ \text{moles}\ O_2}{h} * 100}{\text{MDRYZI}\ \frac{\#\ \text{moles dry gas}}{h}}$$

$$= \frac{\text{MO2ZO * 100}}{\text{MDRYZI}}$$, the percent oxygen by volume in the scrubber inlet gas on a dry basis.

$$\text{PH2OZI} = \frac{\text{MH2OZ*}\ \frac{\#\ \text{moles}\ H_2O}{h} * 100}{\text{MTOTZI}\ \frac{\#\ \text{moles gas}}{h}}$$

$$= \frac{\text{MH2OZI * 100}}{\text{MTOTZI}}$$, the percent water by volume in the scrubber inlet gas.

$$\text{EXO2} = \frac{\text{MO2ZO}\ \frac{\#\ \text{moles}\ O_2}{h}}{\text{MO2RQ}\ \frac{\#\ \text{moles}\ O_2}{h} + \text{MO2X}\ \frac{\#\ \text{moles}\ O_2\ \text{in coal}}{h}} * 100$$

$$= \frac{\text{100 * MO2ZO}}{[\text{MO2RQ + MO2X}]}$$, the percent excess air.

Rapid Estimation of Flue Gas Flows from Continuous Measurements

Curves have been developed that can be used to estimate the flue gas flow from readily available measurements of coal rate and of flue gas CO_2 or O_2 concentration (Figures D-1 and D-2). These curves were developed for a coal composition which is the mean of the coal samples representing the nine Baseline II tests. The CO_2, O_2, and coal rate measurements are available as one hour averages from the DAS.

Using Figure D-2 if the coal rate is 11.35 kg/s (90,000 lb/h) and if there is 12% CO_2 in the flue gas, the flow into the scrubber will be 180.3 m^3/s (382,000 ACFM) at 148.9°C (300°F) and 101.3 kPa (29.92" Hg).

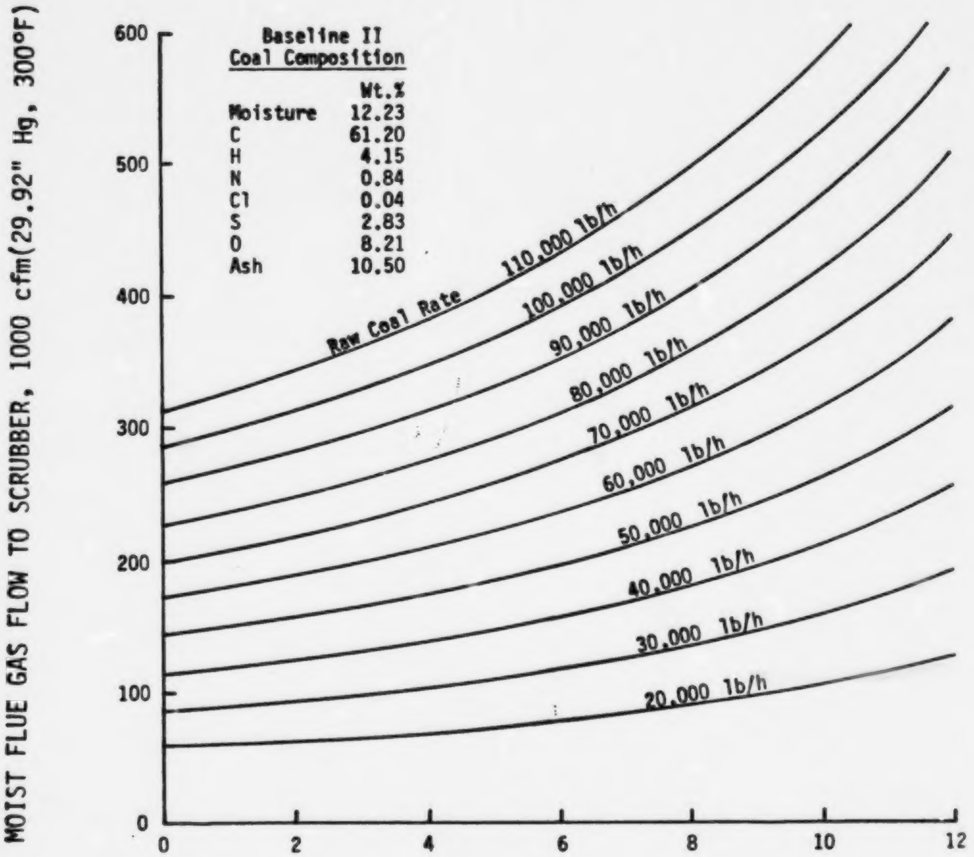
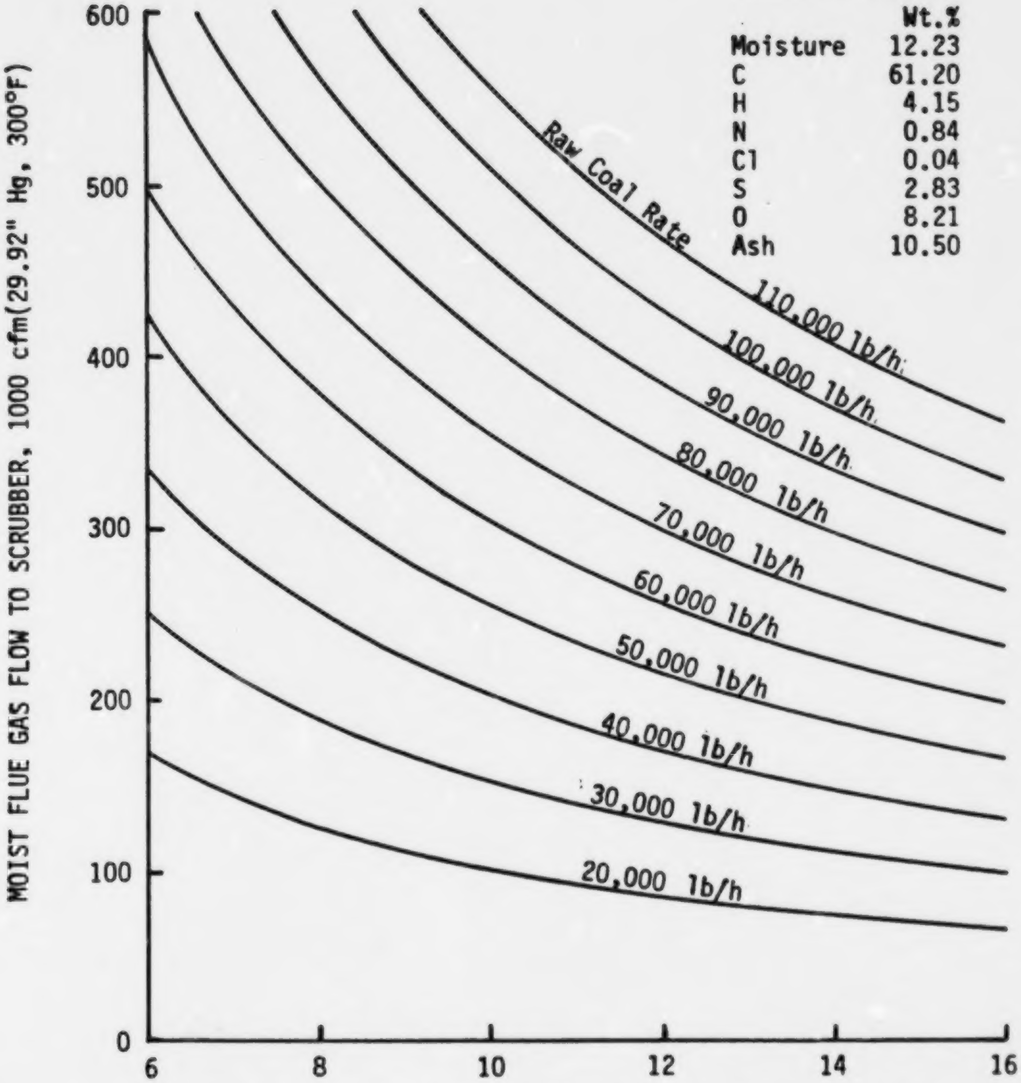


Figure D-1. Percent O_2 in flue gas by volume, dry basis.



Baseline II
Coal Composition

	Wt.%
Moisture	12.23
C	61.20
H	4.15
N	0.84
Cl	0.04
S	2.83
O	8.21
Ash	10.50

Figure D-2. Percent CO_2 in flue gas by volume, dry basis.

APPENDIX E

METHOD FOR CALCULATING SULFUR BALANCE

APPENDIX E

METHOD FOR CALCULATING SULFUR BALANCE

Several different sets and subsets of the fate of the available sulfur were determinable:

(a) Sulfur balance across boiler alone.

(b) Sulfur balance across boiler and FGD plant by
  - determination of FGD plant inlet sulfur from coal
  - determination of FGD plant inlet sulfur from flue gas analysis

(c) Sulfur balance across FGD only by
  - determination of FGD plant inlet sulfur from coal
  - determination of FGD plant inlet sulfur from flue gas analysis

(d) Repeat any of the above for either Mode 2 operation or all mode operation.

Procedure

The minimum frequency of determination is one month due to the availabilities of soda ash consumption and sulfur production. Total quantities for a 30-day period were determined with all sulfur forms reported as sulfur. Quantities were determined for each FGD operating mode (Modes 2, 3, 4). The steps required are as follows.

1. Determine total coal burned for each mode and total.
2. Sulfur in coal. Coal analyses were done no more often than every sixth day. The computer calculated sulfur in coal for each hour and summed. These data were distributed to each operating mode.

   Total = Mode 2 + 0.75 Mode 3 + Mode 4
3. Ash in coal. Determine ash in coal for each mode and total.
4. Sulfur in ash. Sulfur = ash x 0.0064.
5. Sulfur in flue gas, coal basis; sulfur = sulfur in coal-sulfur in ash.
6. Sulfur in flue gas, flue gas analysis basis. The computer calculated inlet rate of SO_2 for each hour that SO_2 and flue gas flow data were available and summed. Hourly records were used to factor up to the total quantity of sulfur fed:

$$\text{sulfur in (Mode 2)} = \frac{\text{total } SO_2 \text{ measured}}{2} \times \frac{\text{total records of mode}}{\text{total measurement records}}$$

   sulfur in (total) = sulfur in, Mode 2 + 0.75 sulfur in, Mode 3 + sulfur in, Mode 4.
7. Sulfur in fly ash purge.

   Assumptions: purge contains 0.13 percent sulfur by weight.

   Purge rate is 70.9 gpm.

   Sulfur = 46.16 R

   Where:

   R = total records of mode or period (use 0.75 x Mode 3 records).

8. Sulfur removed. This is the total sulfur removed, including sulfur in the fly ash purge.

   (a) Determine average percent removed for each mode.

   (b) Sulfur removed = sulfur in flue gas x % removal.

9. Sulfur out stack.

   Sulfur = sulfur in flue gas - sulfur removed - sulfur in fly ash purge.

10. Sulfur to recovery.

    Sulfur = sulfur removal - sulfur in fly ash purge.

11. Sulfur in purge. It is assumed that a mole of sulfur ends up in the purge for every 2 moles of sodium in the soda ash makeup. The monthly consumption of soda ash is used:

$$\frac{\text{lb. sulfur}}{\text{lb soda ash}} = \frac{\text{2 mole Na}}{\text{mole } Na_2CO_3} \times \frac{\text{mole } Na_2CO_3}{\text{105.99 lb. } Na_2CO_3} \times \frac{\text{1 mole S}}{\text{2 mole Na}}$$

$$\times \frac{\text{32 lbs}}{\text{mole S}} = \frac{\text{0.3019 lb S}}{\text{lb soda ash}}$$

    sulfur in purge (total) = 0.3019 x lb. soda ash.

$$\text{sulfur in purge (Mode 2)} = \frac{\text{0.3019 x lb soda ash x Mode 2 records}}{\text{records (Mode 2 +0.75 Mode 3 + Mode 4)}}$$

12. Sulfur product.

    Sulfur product (total) = production for month

    Sulfur product (Mode 2) = production for month x

$$\frac{\text{Mode 2 months}}{\text{records (Mode 2 + 0.75 mode 3)}}$$

13. Sulfur bypassed, for total operating period only. During Mode 3, it is assumed that 75% of the inlet flue gas is taken by the FGD plant and that 25% is bypassed.

During Mode 4, part of the sulfur to recovery goes to purge and the rest is bypassed. Total sulfur bypassed is the sum of these two quantities.

14. Sulfur balance, boiler + FGD plant.

Sulfur in	Sulfur out
In coal	With ash
	With fly ash purge
	Out stack
	With Purge
	With product
	Bypassed

$$\text{Percent closure} = \frac{\text{sulfur out}}{\text{sulfur in}} \times 100$$

15. Sulfur balance, FGD plant

Sulfur in	Sulfur out
Flue gas in	With fly ash purge
	Out stack
	With purge
	With product
	Bypassed

$$\text{Percent closure} = \frac{\text{sulfur out}}{\text{sulfur in}} \quad 100$$

APPENDIX F

PURGE DRYER TEST OBSERVATIONS

APPENDIX F

PURGE DRYER TEST OBSERVATIONS

The test contractor observed tests performed by Davy McKee. The dryer was tested for approximately one week during May 1979. The tests verified that the dryer did not have heat duty design capacity. The purpose of the test was to demonstrate the heat capacity of the dryer with a water feed onto a sodium sulfate bed in the dryer. If the maximum rate could be reached with water, the dryer was then to be tested with the sodium sulfate solution recovered from the purge solution.

The maximum dryer capacity achieved during the test was approximately 290,000 J/s (1.0 million Btu/h), which is 66% of the design heat duty of 463,000 J/s (1.58 million Btu/h).[14] Capacities of 59-66% of design were maintained for 2½ days. However, after operating for 2½ days at 59-66%, the motor tripped out several times because of an amperage overload. When the motor tripped out, a buildup of solids was observed at the discharge end of the dryer. To prevent overloading of the motor, the water feed rate was reduced; thus during the last 2½ days of the test, the dryer capacity dropped to 45-50% of design. At this point, further testing was abandoned.

Plots of the solution feed rate and dryer heat duty are given in Figure F-1, motor amperage range in Figure F-2, and dryer capacity in Figure F-3. Table F-1 gives the calculated heat duty, and Tables F-2 and F-3 show the test data collected by Davy Powergas.

Purge Dryer Test Results

The NIPSCO purge dryer removes most of the moisture from the sodium sulfate crystallized from the absorber purge solution. The original design moisture content of the sulfate purge solids was 0.2% by weight, but a moisture content of 0.05% or less is desirable. Since startup, Allied reported that the purge dryer would not achieve design capacity. The dryer had been fouling, thus reducing the heat transfer efficiency. Allied was unable to find a set of operating conditions to prevent fouling. Moreover, the only way to remove the fouling was to shut down the dryer, remove the bed, and manually scrape and chip off the coating.

Because of these operating problems, the dryer was tested by Davy McKee for approximately one week during the period of May 25-31, 1979. Normally, the dryer feed is a suspension of salt crystals in water. However, during the tests, water was used in place of the salt-bearing feed, and the dryer was filled to the approximate normal bed depth with dry sodium sulfate. Four spray nozzles were installed in the dryer, and the total water fed to all of the nozzles was measured with a rotameter. The flow was varied by changing the number of nozzles or by altering the water pressure During the test, Davy recorded the bed temperature, the pressure and temperature of the steam, the motor amps, and the temperature, pressure, and flow rate of the water. These data, recorded at two-hour intervals, are listed in Table F-1. The moisture content of the sodium sulfate product is listed in Table F-2.

From the water feed rate, the heating capacity of the dryer was calculated; the results are summarized in Table F-1. The dryer capacities in this table were calculated from the heat of vaporization of the incoming water. Sensible heat changes in the water, water vapor, and leakage air were neglected. Sensible heat changes might increase the dryer capacities listed in Table F-1 by up to 10%. Also listed in Table F-1 is the ratio of the experimental heat duty to the dryer design heat duty, expressed as a percentage. The design heat duty of the dryer is 463,000 J/s (1.58 million Btu/h). During the test, the dryer capacity varied from 42 to 67%, or from 190,000 to 310,000 J/s (0.66 to 1.06 million Btu/h).

The water feed rate to the dryer and the heat duty are plotted in Figure F-1. A feed rate of approximately 20.8 x 10^{-5} m^3/s (3.3 gpm) is required to attain the dryer design heat duty of 463,000 J/s (1.58 million Btu/h). For the first day, the water feed rate was approximately 9.5 x 10^{-5} m^3/s (1.5 gpm), and the dryer heat duty was 200,000 J/s (0.7 million Btu/h). Then, for the next 2½ days, the water feed rate was approximately 13.9 x 10^{-5} m^3/s (2.2 gpm), corresponding to a heat duty of 290,000 J/s (1.0 million Btu/h). Finally, for the last 2½ days of the test, the flow rate dropped to 9.5 x 10^{-5} m^3/s (1.5 gpm) and the heat duty to 230,000 J/s (0.8 million Btu/h).

The motor amp ranges during the dryer test are plotted in Figure F-2. An amperage range is plotted because the upper and lower amperage limits in each two-hour interval were recorded. The amperage increased from 55-59 amps on May 25 to 65-75 amps on May 29. The motor tripped out several times on May 29 because of an amperage overload. At the end of the test on May 31, the amperage range was 62-68 amps.

The dryer capacity as a percentage of the design heat duty is plotted in Figure F-3. For the first day of the test, the dryer operated at 46% of capacity. Then, a capacity of 60-66% was maintained for 2½ days until overloading of the motor became a problem on May 29. When the motor tripped out, a buildup of solids was observed at the discharge end of the dryer. To prevent overloading of the motor, the water feed rate was reduced; thus during the last 2½ days of the test, the dryer capacity dropped to 45-50% of design.

Conclusions

1. The maximum capacity the purge dryer achieved during the test was 66% of the design heat duty.
2. Capacities of 60-66% were maintained for 2½ days.
3. Fouling of the heat transfer surfaces cannot be avoided, even when only water is used as feed.

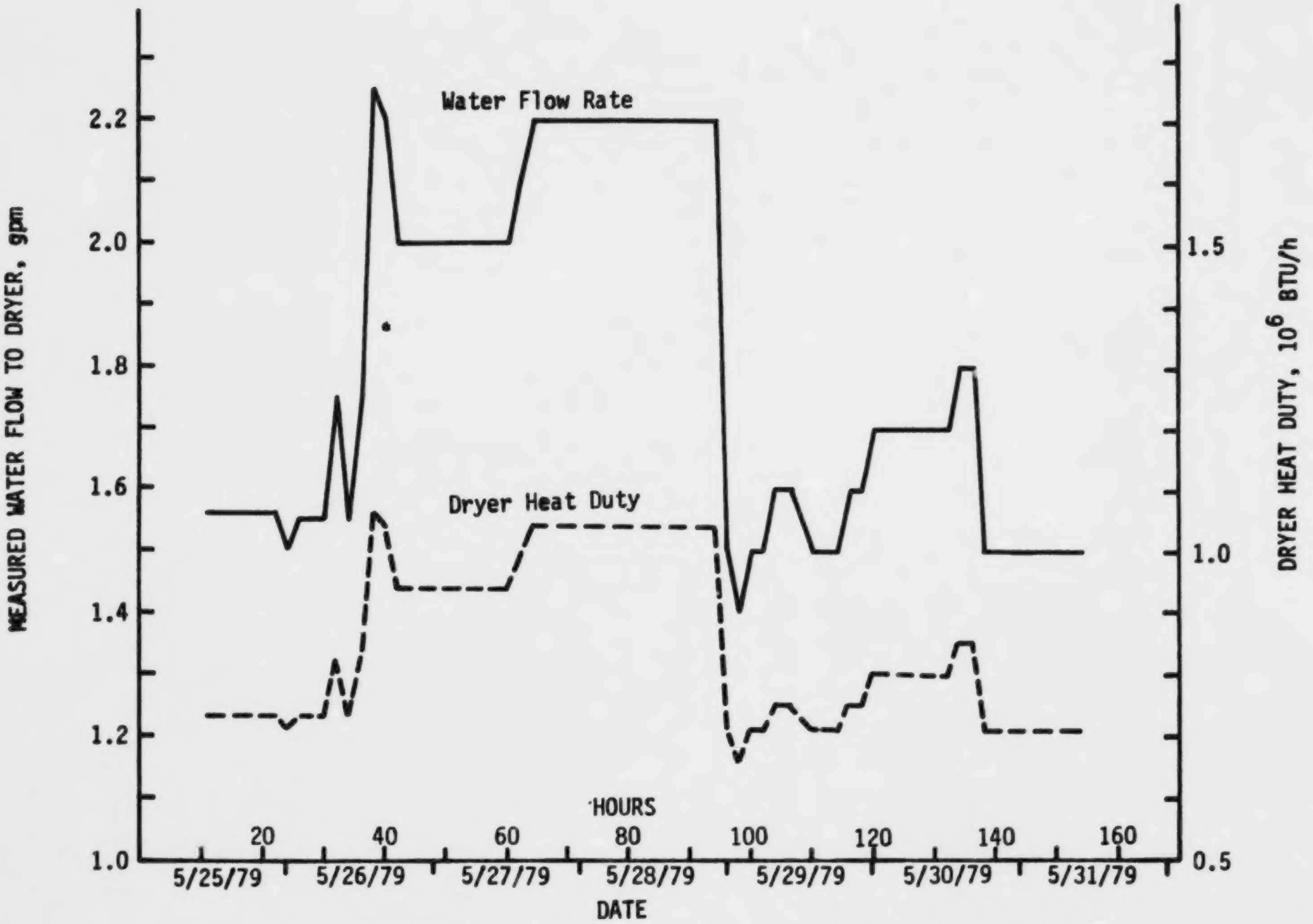


Figure F-1. Solution feed rate and heat duty during purge dryer tests.

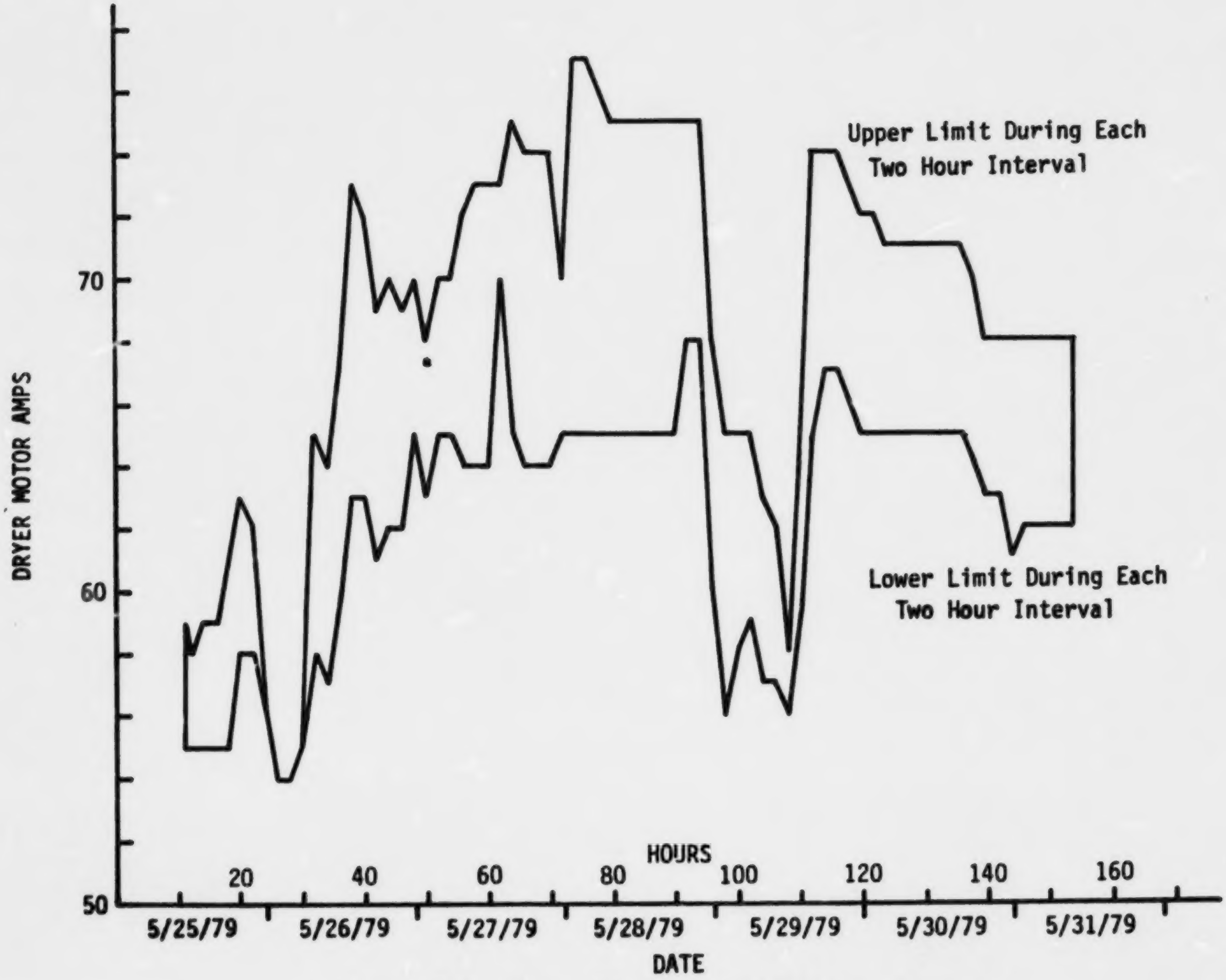
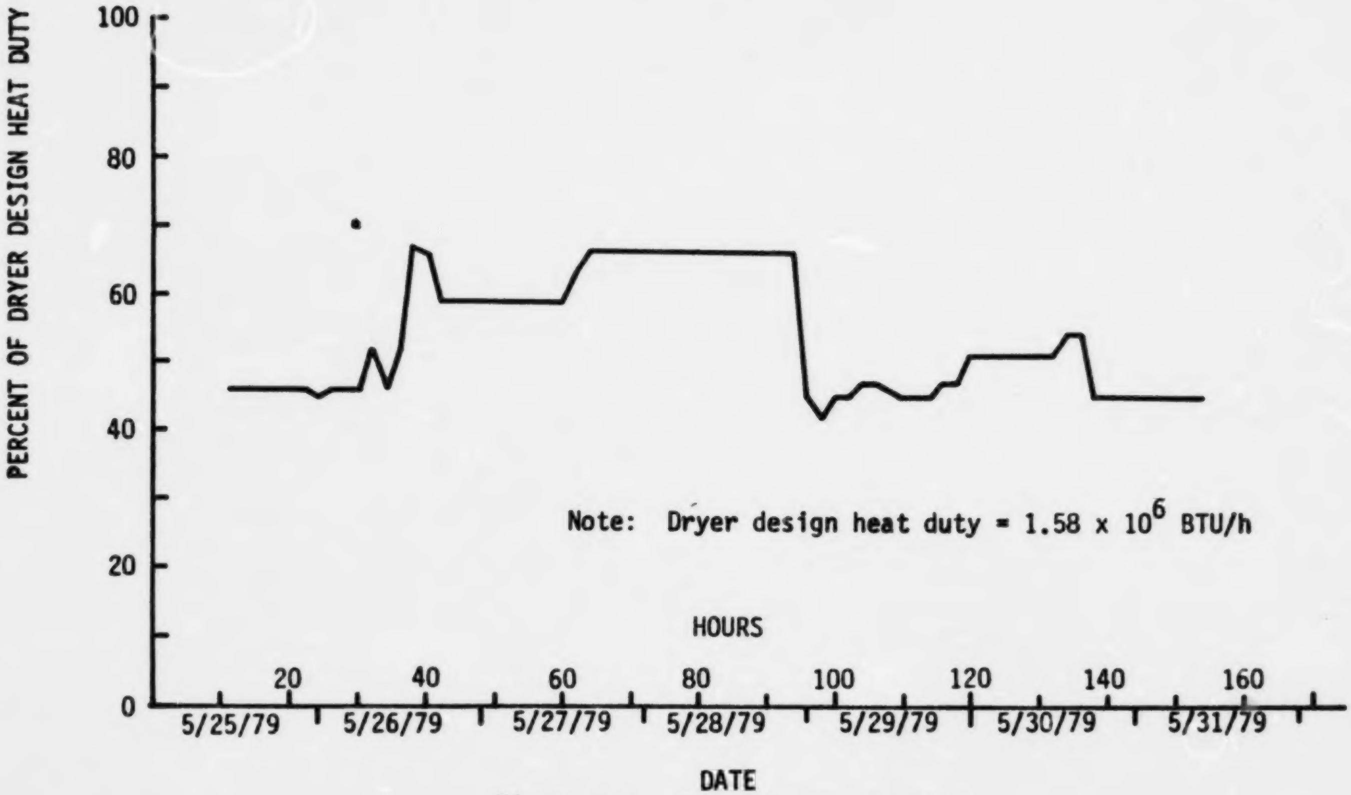


Figure F-2. Motor AMP range during purge dryer tests.



Note: Dryer design heat duty = 1.58×10^6 BTU/h

Figure F-3. Purge dryer capacity.

TABLE F-1A. PURGE DRYER HEAT DUTY - METRIC UNITS

Time	Nozzle	P KPa	T °C	Meas. Flow m^3/s	Heat Duty Based on Meas. Flow J/s	% Dryer Design DUTY*
5/25/79						
11:00 AM	A+D	197	79.4	9.84×10^{-5}	210,000	46
12:00 Noon	"	197	79.4	9.84×10^{-5}	210,000	46
2:00 PM	"	197	80	9.84×10^{-5}	210,000	46
4:00 PM	B+D	197	78.9	9.84×10^{-5}	210,000	46
6:00 PM	A+D	98	80.6	9.84×10^{-5}	210,000	46
8:00 PM	"	98	80	9.84×10^{-5}	210,000	46
10:00 PM	"	98	79.4	9.84×10^{-5}	210,000	46
5/26/79						
12:00 Midnt.	A+D	98	80.6	9.5×10^{-5}	210,000	45
2:00 AM	"	98	71.1	9.78×10^{-5}	210,000	46
4:00 AM	"	98.25	77.2	9.78×10^{-5}	210,000	46
6:00 AM	"	98.25	79.4	9.78×10^{-5}	210,000	46
8:00 AM	"	98	83.3	11.0×10^{-5}	240,000	52
10:00 AM	"	79.3	83.3	9.78×10^{-5}	210,000	46
12:00 Noon	3 nozzles	98	85.6	11.0×10^{-5}	240,000	52
2:00 PM	"	103	87.2	14.2×10^{-5}	310,000	67
4:00 PM	"	88.3	86.1	13.9×10^{-5}	305,000	66
6:00 PM	"	69.6	83.9	12.6×10^{-5}	280,000	59
8:00 PM	"	68.9	83.3	12.6×10^{-5}	280,000	59
10:00 PM	"	67	82.8	12.6×10^{-5}	280,000	59
5/27/79						
12:00 Midnt.	3 nozzles	67.9	81.7	12.6×10^{-5}	280,000	59
2:00 AM	"	67	82.2	12.6×10^{-5}	280,000	59
4:00 AM	"	67.9	82.2	12.6×10^{-5}	280,000	59
6:00 AM	"	67	82.2	12.6×10^{-5}	280,000	59
8:00 AM	"	67	82.2	12.6×10^{-5}	280,000	59
10:00 AM	"	68.9	83.3	12.6×10^{-5}	280,000	59
12:00 Noon	"	75.8	83.3	12.6×10^{-5}	280,000	59
2:00 PM	"	79.3	83.3	13.2×10^{-5}	290,000	63
4:00 PM	"	85.5	83.3	13.9×10^{-5}	305,000	66
6:00 PM	"	85.5	83.9	13.9×10^{-5}	305,000	66
8:00 PM	"	85.5	83.9	13.9×10^{-5}	305,000	66
10:00 PM	"	85.5	82.2	13.9×10^{-5}	305,000	66

(continued)

TABLE F-1A (continued)

Time	Nozzle	P KPa	T ^oC	Meas. Flow m^3/s	Heat Duty Based on Meas. Flow J/s	% Dryer Design DUTY*
5/28/79						
12:00 Midnt.	3 nozzles	84.1	80.6	13.9×10^{-5}	305,000	66
2:00 AM	"	84.1	82.2	13.9×10^{-5}	305,000	66
4:00 AM	"	84.1	82.2	13.9×10^{-5}	305,000	66
6:00 AM	"	84.1	81.7	13.9×10^{-5}	305,000	66
8:00 AM	"	84.1	81.7	13.9×10^{-5}	305,000	66
10:00 AM	"	82.7	83.3	13.9×10^{-5}	305,000	66
12:00 Noon	"	84.8	83.3	13.9×10^{-5}	305,000	66
2:00 PM	"	84.1	82.8	13.9×10^{-5}	305,000	66
4:00 PM	"	84.1	83.3	13.9×10^{-5}	305,000	66
6:00 PM	"	84.1	83.3	13.9×10^{-5}	305,000	66
8:00 PM	"	84.8	82.2	13.9×10^{-5}	305,000	66
10:00 PM	"	98.6	80.6	13.9×10^{-5}	305,000	66
5/29/79						
12:00 Midnt.	3 nozzles	78.6	80	9.5×10^{-5}	210,000	45
2:00 AM	"	80.0	80	8.8×10^{-5}	190,000	42
4:00 AM	"	79.3	81.1	9.5×10^{-5}	210,000	45
6:00 AM	"	79.3	80.6	9.5×10^{-5}	210,000	45
8:00 AM	"	80.0	80.6	10.1×10^{-5}	220,000	47
10:00 AM	"	80.0	81.7	10.1×10^{-5}	220,000	47
12:00 Noon	"	36.5	75.6	9.8×10^{-5}	210,000	46
2:00 PM	"	39.3	75.6	9.5×10^{-5}	210,000	45
4:00 PM	4 nozzles	39.3	80	9.5×10^{-5}	210,000	45
6:00 PM	"	42.1	80.6	9.5×10^{-5}	210,000	45
8:00 PM	"	41.7	80.6	10.1×10^{-5}	220,000	47
10:00 PM	"	41.4	80	10.1×10^{-5}	220,000	47
5/30/79						
12:00 Midnt.	4 nozzles	43.4	80	10.7×10^{-5}	230,000	51
2:00 AM	"	43.4	78.9	10.7×10^{-5}	230,000	51
4:00 AM	"	43.4	79.4	10.7×10^{-5}	230,000	51
6:00 AM	"	43.4	81.1	10.7×10^{-5}	230,000	51
8:00 AM	"	43.4	81.1	10.7×10^{-5}	230,000	51
10:00 AM	"	49.6	82.2	10.7×10^{-5}	230,000	51
12:00 Noon	"	49.6	82.8	10.7×10^{-5}	230,000	51
2:00 PM	"	56.5	82.2	11.4×10^{-5}	250,000	54
4:00 PM	"	57.2	82.2	11.4×10^{-5}	250,000	54

(continued)

TABLE F-1A (Concluded)

Time	Nozzle	P KPa	T ^{o}C	Meas. Flow m^3/s	Heat Duty Based on Meas. Flow J/s	% Dryer Design DUTY*
5/30/79						
6:00 PM	3 nozzles	57.9	80	9.5×10^{-5}	210,000	45
8:00 PM	"	57.9	81.7	9.5×10^{-5}	210,000	45
10:00 PM	"	27.9	82.2	9.5×10^{-5}	210,000	45
5/31/79						
12:00 Midnt.	3 nozzles	57.9	82.2	9.5×10^{-5}	210,000	
2:00 AM	"	57.9	81.1	9.5×10^{-5}	210,000	45
4:00 AM	"	57.9	80.6	9.5×10^{-5}	210,000	45
6:00 AM	"	57.9	80	9.5×10^{-5}	210,000	45
8:00 AM	"	57.9	79.4	9.5×10^{-5}	210,000	45
10:00 AM	"	57.9	80	9.5×10^{-5}	210,000	45

*NOTE: Dryer design heat duty = 1.58×10^6 Btu/h(2)

TABLE F-1B. PURGE DRYER HEAT DUTY - ENGLISH UNITS

Time	Nozzle	P (PSI)	T (°F)	Meas. Flow (GPM)	Heat Duty Based on Meas. Flow (Million Btu/h)	% Dryer Design Duty*
5/25/79						
11:00 AM	A+D	28.6	175	1.56	0.73	46
12:00 Noon	"	28.6	175	1.56	0.73	46
2:00 PM	"	28.6	176	1.56	0.73	46
4:00 PM	B+D	28.6	174	1.56	0.73	46
6:00 PM	A+D	14.2	177	1.56	0.73	46
8:00 PM	"	14.2	176	1.56	0.73	46
10:00 PM	"	14.2	175	1.56	0.73	56
5/26/79						
12:00 Midnt.	A+D	14.2	177	1.5	0.71	45
2:00 AM	"	14.2	160	1.55	0.73	46
4:00 AM	"	14.25	171	1.55	0.73	46
6:00 AM	"	14.25	175	1.55	0.73	46
8:00 AM	"	14.2	182	1.75	0.82	52
10:00 AM	"	11.5	182	1.55	0.73	46
12:00 Noon	3 nozzles	14.2	186	1.75	0.82	52
2:00 PM	"	15	189	2.25	1.06	67
4:00 PM	"	12.8	187	2.2	1.04	66
6:00 PM	"	10.1	183	2.0	0.94	59
8:00 PM	"	10.0	182	2.0	0.94	59
10:00 PM	"	9.8	181	2.0	0.94	59
5/27/79						
12:00 Midnt.	3 nozzles	9.85	179	2.0	0.94	59
2:00 AM	"	9.8	180	2.0	0.94	59
4:00 AM	"	9.85	180	2.0	0.94	59
6:00 AM	"	9.8	180	2.0	0.94	59
8:00 AM	"	9.8	180	2.0	0.94	59
10:00 AM	"	10.0	182	2.0	0.94	59
12:00 Noon	"	11.0	182	2.0	0.94	59
2:00 PM	"	11.5	182	2.1	0.99	63
4:00 PM	"	12.4	182	2.2	1.04	66
6:00 PM	"	12.4	183	2.2	1.04	66
8:00 PM	"	12.4	183	2.2	1.04	66
10:00 PM	"	12.4	180	2.2	1.04	66

(continued)

TABLE F-1B (continued)

Time	Nozzle	P (PSI)	T (°F)	Meas. Flow (GPM)	Heat Duty Based On Meas. Flow (Million Btu/h)	% Dryer Design Duty*
5/28/79						
12:00 Midnt.	3 nozzles	12.2	177	2.2	1.04	66
2:00 AM	"	12.2	180	2.2	1.04	66
4:00 AM	"	12.2	180	2.2	1.04	66
6:00 AM	"					
8:00 AM	"	12.2	179	2.2	1.04	66
10:00 AM	"	12.0	182	2.2	1.04	66
12:00 Noon	"	12.3	182	2.2	1.04	66
2:00 PM	"	12.2	181	2.2	1.04	66
4:00 PM	"	12.2	182	2.2	1.04	66
6:00 PM	"	12.2	182	2.2	1.04	66
8:00 PM	"	12.3	181	2.2	1.04	66
10:00 PM	"	14.3	177	2.2	1.04	66
5/29/79						
12:00 Midnt.	3 nozzles	11.4	176	1.5	0.71	45
2:00 AM	"	11.6	176	1.4	0.66	42
4:00 AM	"	11.5	178	1.5	0.71	45
6:00 AM	"	11.5	177	1.5	0.71	45
8:00 AM	"	11.6	177	1.6	0.75	47
10:00 AM	"	11.6	179	1.6	0.75	47
12:00 Noon	"	5.3	168	1.55	0.73	46
2:00 PM	"	5.7	168	1.5	0.71	45
4:00 PM	4 nozzles	5.7	176	1.5	0.71	45
6:00 PM	"	6.1	177	1.5	0.71	45
8:00 PM	"	6.05	177	1.6	0.75	47
10:00 PM	"	6.0	176	1.6	0.75	47
5/30/79						
12:00 Midnt.	4 nozzles	6.3	176	1.7	0.80	51
2:00 AM	"	6.3	174	1.7	0.80	51
4:00 AM	"	6.3	175	1.7	0.80	51
6:00 AM	"	6.3	178	1.7	0.80	51
8:00 AM	"	6.3	178	1.7	0.80	51
10:00 AM	"	7.2	180	1.7	0.80	51
12:00 Noon	"	7.2	181	1.7	0.80	51
2:00 PM	"	8.2	180	1.8	0.85	54
4:00 PM	"	8.3	180	1.8	0.85	54
5/30/79						
6:00 PM	3 nozzles	8.4	176	1.5	0.71	45
8:00 PM	"	8.4	179	1.5	0.71	45
10:00 PM	"	8.4	180	1.5	0.71	45

(continued)

TABLE F-1B (concluded)

Time	Nozzle	P (PSI)	T (^oF)	Meas. Flow (GPM)	Heat Duty Based On Meas. Flow (Million Btu/h)	% Dryer Design Duty*
5/31/79						
12:00 Midnt.	3 nozzles	8.4	180	1.5	0.71	45
2:00 AM	"	8.4	178	1.5	0.71	45
4:00 AM	"	8.4	177	1.5	0.71	45
6:00 AM	"	8.4	176	1.5	0.71	45
8:00 AM	"	8.4	175	1.5	0.71	45
10:00 AM	"	8.4	176	1.5	0.71	45

*NOTE: Dryer design heat duty = 1.58×10^6 Btu/h(2)

TABLE F-2. PURGE DRYER PERFORMANCE

Date	Time	Wt. % H_2O in Solids at Discharge
05/29/79	8:00 AM	0.065
	1:21 PM	0.02
	5:00 PM	0.052
	10:00 PM	0.056
05/30/79	8:00 AM	0.026
	12:45 PM	0.160 Bed Temp = 157.8°C (316°F)
	3:00 PM	0.120 Bed Temp = 153.9°C (309°F)
05/31/79	8:00 AM	0.021 Bed Temp = 164.4°C (328°F)

TABLE F-3A. PURGE DRYER TEST DATA - METRIC UNITS
DATE: 5/25/79

Time	BED	STEAM		MOTOR AMPS	FEED			Comments
	TEMP °C	TEMP °C	PRESS KPa		TEMP °C	RATE m^3/s	PRESS KPa	
11:00 AM	157.2	178.3	931	55-59	79.4	9.84×10^{-5}	197	A & D on.
12:00 Noon	151.7	181.1	979	55-58	79.4	9.84×10^{-5}	197	
3:00 PM	151.1	181.7	1007	55-59	80.0	9.84×10^{-5}	197	
4:00 PM	152.8	183.9	1062	55-59	78.9	9.84×10^{-5}	197	"B"-spray on. "A" off.
6:00 PM	154.4	182.8	1048	55-61	80.6	9.84×10^{-5}	97.9	"A" spray back on.
8:00 PM	152.8	183.3	1048	58-63	80.0	9.84×10^{-5}	97.9	
10:00 PM	152.8	183.3	1048	58-62	79.4	9.84×10^{-5}	97.9	

(continued)

TABLE F-3A (continued)
DATE: 5/26/79

Time	BED	STEAM		MOTOR AMPS	FEED			Comments
	TEMP °C	TEMP °C	PRESS KPa		TEMP °C	RATE m^3/s	PRESS KPa	
12:00 Midnt.	152.2	183.3	1041	56	80.6	9.5×10^{-5}	97.9	Condensate hose feeding dry boiled at 12:30 AM. Repaired by 1:30 AM.
2:00 AM	165.6	185.6	1076	54	71.1	9.78×10^{-5}	97.9	Due to feed hose breaking.
4:00 AM	153.9	182.2	1041	54	77.2	9.78×10^{-5}	98.25	
6:00 AM	151.7	182.2	1027	55	79.4	9.78×10^{-5}	98.25	
8:00 AM	153.9	183.9	1062	58-65	83.3	11.0×10^{-5}	97.9	Moisture content .09%.
10:00 AM	157.8	152.2	1020	57-64	83.3	9.78×10^{-5}	79.3	Changed to 75% flow after this reading.
12:00 Noon	150.5	183.3	1062	60-67	85.6	11.0×10^{-5}	97.9	Had to adjust flow due to upset in steam line back to 50%. Changed flow to 75% at 12:00. At this time, bed temp drops rapidly. See P. Lechner's note.
2:00 PM	136.1	183.9	1062	63-73	87.2	14.2×10^{-5}	103	
4:00 PM	140	183.9	1069	63-72	86.1	13.9×10^{-5}	82.3	Steam blowing from dryer.
6:00 PM	145.6	183.3	1062	61-69	83.9	12.6×10^{-5}	69.6	
8:00 PM	143.9	183.9	1062	62-70	83.3	12.6×10^{-5}	68.9	
10:00 PM	145.6	183.9	1062	62-69	82.8	12.6×10^{-5}	57.6	

(continued)

TABLE F-3A (continued)
DATE: 5/27/79

Time	BED	STEAM		MOTOR AMPS	FEED			Comments
	TEMP °C	TEMP °C	PRESS KPa		TEMP °C	RATE m^3/s	PRESS KPa	
12:00 Midn.	144.4	183.3	1048	65-70	81.7	12.6×10^{-5}	67.9	
2:00 AM	143.3	182.8	1055	63-68	82.2	12.6×10^{-5}	67.6	
4:00 AM	143.3	183.3	1048	65-70	82.2	12.6×10^{-5}	67.9	
6:00 AM	145.0	183.3	1041	65-70	82.2	12.6×10^{-5}	67.6	
8:00 AM	145.6	183.3	1055	64-72	82.2	12.6×10^{-5}	67.6	Flow increased incrementally. Moisture = .15%.
10:00 AM	145.6	183.9	1055	64-73	83.3	12.6×10^{-5}	68.9	Same as above.
12:00 Noon	143.9	183.9	1069	64-73	83.3	12.6×10^{-5}	75.8	Same as above.
2:00 PM	143.3	184.2	1076	70-73	83.3	13.2×10^{-5}	79.3	Same as above.
4:00 PM	140.6	183.9	1069	65-75	83.3	13.9×10^{-5}	85.5	Same as above.
6:00 PM	140.6	183.9	1062	64-74	83.9	13.9×10^{-5}	85.5	Flow reduced incrementally by decreasing pressure by 1.4 m^3/s.
8:00 PM	141.1	183.9	1062	64-74	83.9	13.9×10^{-5}	85.5	
10:00 PM	141.7	183.9	1062	64-74	82.2	13.9×10^{-5}	85.5	

(continued)

TABLE F-3A (continued)
DATE: 5/28/79

Time	BED	STEAM		MOTOR AMPS	FEED			Comments
	TEMP °C	TEMP °C	PRESS KPa		TEMP °C	RATE m^3/s	PRESS KPa	
12:00 Midnt.	140.6	183.9	1055	65-70	80.6	13.9×10^{-5}	84.1	
2:00 AM	140.6	183.3	1048	65-77	82.2	13.9×10^{-5}	84.1	
4:00 AM	141.1	183.3		65-77	82.2	13.9×10^{-5}	84.1	
6:00 AM								
8:00 AM	140.0	183.3	1048	65-75	81.7	13.9×10^{-5}	84.1	
10:00 AM	141.1	183.9	1062	65-75	83.3	13.9×10^{-5}	82.7	
12:00 Noon	140.6	182.8	1048	65-75	83.3	13.9×10^{-5}	84.8	
2:00 PM	140.6	183.3	1055	65-75	82.8	13.9×10^{-5}	84.1	
4:00 PM	141.7	183.9	1069	67-75	83.3	13.9×10^{-5}	84.1	
6:00 PM	142.2	138.9	1069	65-75	83.3	13.9×10^{-5}	84.1	
8:00 PM	142.8	183.9	1062	68-75+	82.8	13.9×10^{-5}	84.8	Dryer K.O. at 7:10 PM.
10:00 PM	143.3	184.4	1069	68-75+	80.6	11.0×10^{-5}	98.6	Dryer K.O. at 10:04 PM.

(continued)

TABLE F-3A (continued)
DATE: 5/29/79

Time	BED	STEAM		MOTOR AMPS	FEED			Comments
	TEMP $^{\circ}$C	TEMP $^{\circ}$C	PRESS KPa		TEMP $^{\circ}$C	RATE m^3/s	PRESS KPa	
12:00 Midnt.	154.4	183.3	1048	60-68	80.0	9.5×10^{-5}	78.6	
2:00 AM	155.6	183.3	1034	56-65	80.0	8.8×10^{-5}	80.0	
4:00 AM	154.4	183.3	1048	58-65	81.1	9.5×10^{-5}	79.3	
6:00 AM	154.4	183.3	1055	59-65	80.6	9.5×10^{-5}	79.3	
8:00 AM	161.1	183.9	1076	57-63	80.6	10.1×10^{-5}	80.0	
10:00 AM	161.1	183.9	1048	57-62	81.7	10.1×10^{-5}	80.0	Dryer shut down for inspection.
12:00 Noon	165.6	184.4	1048	56-58	75.6	9.8×10^{-5}	36.5	11:00, dryer restored; at 1:00 moisture .02%.
2:00 PM	178.9	185	1062	59-64	75.6	9.5×10^{-5}	39.3	3 sprays.
4:00 PM	160.6	183.9	1076	65-74	80.0	9.5×10^{-5}	39.3	4 sprays.
6:00 PM	161.1	183.9	1082	67-74	80.6	9.5×10^{-5}	42.1	5:00 took samples.
8:00 PM	161.7	183.9	1076	67-74	80.6	10.1×10^{-5}	41.7	
10:00 PM	161.7	183.9	1076	66-73	80.0	10.1×10^{-5}	41.4	Took sample for moisture dryer steaming.

(continued)

TABLE F-3A (continued)
DATE: 5/30/79

Time	BED	STEAM		MOTOR AMPS	FEED			Comments
	TEMP °C	TEMP °C	PRESS KPa		TEMP °C	RATE m^3/s	PRESS KPa	
12:00 Midnt.	154.4	183.3	1034	65-73	80.0	10.7×10^{-5}	43.4	
2:00 AM	154.4	183.3	1034	65-72	78.9	10.7×10^{-5}	43.4	
4:00 AM	155.6	183.3	1048	65-71	79.4	10.7×10^{-5}	43.4	
6:00 AM	155.6	183.3	1048	65-71	81.1	10.7×10^{-5}	43.4	
8:00 AM	160.0	183.3	1048	65-71	81.1	10.7×10^{-5}	43.4	
10:00 AM	158.9	183.3	1048	65-71	82.2	10.7×10^{-5}	49.6	
12:00 Noon	158.3	183.3	1048	65-71	82.8	10.7×10^{-5}	49.6	
2:00 PM	155.0	183.3	1048	65-71	82.2	11.4×10^{-5}	49.6	
4:00 PM	153.3	183.3	1062	65-71	82.2	11.4×10^{-5}	57.2	
6:00 PM	163.9	183.3	1062	64-70	80.0	9.5×10^{-5}	57.9	Cut back to 3 spaces at 5:00.
8:00 PM	165.0	183.9	1076	63-68	81.7	9.5×10^{-5}	57.9	
10:00 PM	165.0	183.3	1069	63-68	82.2	9.5×10^{-5}	57.9	

(continued)

TABLE F-3A (continued)
DATE: 5/31/79

Time	BED	STEAM		MOTOR AMPS	FEED			Comments
	TEMP °C	TEMP °C	PRESS KPa		TEMP °C	RATE m^3/s	PRESS KPa	
12:00 Midnt.	165.0	183.3	1062	61-68	82.2	9.5×10^{-5}	58.0	
2:00 AM	165.0	183.3	1062	62-68	81.1	9.5×10^{-5}	55.2	
4:00 AM	165.6	183.9	1069	62-68	80.6	9.5×10^{-5}	58.0	
6:00 AM	165.0	183.3	1069	62-68	80.0	9.5×10^{-5}	58.0	
8:00 AM	164.4	183.3	1048	62-68	79.4	9.5×10^{-5}	58.0	
10:00 AM	165.0	183.3	1048	62-68	80.0	9.5×10^{-5}	58.0	
12:00 Noon								
2:00 PM								
4:00 PM								

TABLE F-3B. PURGE DRYER TEST DATA - ENGLISH UNITS
DATE: 5/25/79

Time	BED	STEAM		MOTOR	FEED			Comments
	TEMP °F	TEMP °F	PRESS PSIG	AMPS	TEMP °F	RATE GPM	PRESS PSI	
11:00 AM	315	353	135	55-59	175	1.56	28.6	A & D on.
12:00 Noon	305	358	142	55-58	175	1.56	28.6	
3:00 PM	304	359	146	55-59	176	1.56	28.6	
4:00 PM	307	363	154	55-59	174	1.56	28.6	"B"-spray on. "A" off.
6:00 PM	310	361	152	55-61	177	1.56	14.2	"A" spray back on.
8:00 PM	307	362	152	58-63	176	1.56	14.2	
10:00 PM	307	362	152	58-62	175	1.56	14.2	

(continued)

TABLE F-3B (continued)
DATE: 5/26/79

	BED	STEAM			FEED			
Time	TEMP ^{o}F	TEMP ^{o}F	PRESS PSIG	MOTOR AMPS	TEMP ^{o}F	RATE GPM	PRESS PSI	Comments
12:00 Midnt.	306	362	151	56	177	1.5	14.2	Condensate hose feeding dry boiled at 12:30 AM. Repaired by 1:30 AM.
2:00 AM	330	366	156	54	160	1.55	14.2	Due to feed hose breaking.
4:00 AM	309	360	151	54	171	1.55	14.25	
6:00 AM	305	360	149	55	175	1.55	14.25	
8:00 AM	309	363	154	58-65	182	1.75	14.2	Moisture content .09%.
10:00 AM	316	360	148	57-64	182	1.55	11.5	Changed to 75% flow after this reading.
12:00 Noon	303	362	154	60-67	186	1.75	14.2	Had to adjust flow due to upset in steam line back to 50%. Changed flow to 75% at 12:00. At this time, bed temp drops rapidly. See P. Lechner's note.
2:00 PM	277	363	154	63-73	189	2.25	15.0	
4:00 PM	284	363	155	63072	187	2.2	12.8	
6:00 PM	294	362	154	61-69	183	2.0	10.1	Steam blowing from dryer.
8:00 PM	291	363	154	62-70	182	2.0	10.0	
10:00 PM	294	363	154	62-69	181	2.0	9.8	

(continued)

TABLE F-3B (continued)
DATE: 5/27/79

Time	BED	STEAM		MOTOR	FEED			Comments
	TEMP °F	TEMP °F	PRESS PSIG	AMPS	TEMP °F	RATE GPM	PRESS PSI	
12:00 Midnt.	292	362	152	65-70	179	2.0	9.85	
2:00 AM	290	361	153	63-68	180	2.0	9.8	
4:00 AM	290	362	152	65-70	180	2.0	9.85	
6:00 AM	293	362	151	65-70	180	2.0	9.8	
8:00 AM	294	362	153	64-72	180	2.0	9.8	Flow increased incrementally. Moisture = .15%.
10:00 AM	295	363	153	64-73	182	2.0	10.0	Same as above.
12:00 Noon	291	363	155	64-73	182	2.0	11.0	Same as above.
2:00 PM	290	363.5	156	70-73	182	2.1	11.5	Same as above.
4:00 PM	285	363	155	65-75	182	2.2	12.4	Same as above.
6:00 PM	285	363	154	64-74	182	2.2	12.4	Flow reduced incrementally by decreasing pressure by 1.4 m^3/s (0.2 psig).
8:00 PM	286	363	154	64-74	182	2.2	12.4	
10:00 PM	287	363	154	64-74	180	2.2	12.4	

(continued)

TABLE F-3B (continued)
DATE: 5/28/79

Time	BED	STEAM		MOTOR AMPS	FEED			Comments
	TEMP °F	TEMP °F	PRESS PSIG		TEMP °F	RATE GPM	PRESS PSI	
12:00 Midn.	285	363	153	65-70	177	2.2	12.2	
2:00 AM	285	362	152	65-77	180	2.2	12.2	
4:00 AM	286	362		65-77	180	2.2	12.2	
6:00 AM								
8:00 AM	284	362	152	65-75	179	2.2	12.2	
10:00 AM	286	363	154	65-75	182	2.2	12.0	
12:00 Noon	285	361	152	65-75	182	2.2	12.3	
2:00 PM	285	362	153	65-75	181	2.2	12.2	
4:00 PM	287	363	155	67-75	182	2.2	12.2	
6:00 PM	288	363	155	65-75	182	2.2	12.2	
8:00 PM	289	363	154	68-75+	181	2.2	12.3	Dryer K.O. at 7:10 PM.
10:00 PM	290	364	155	68-75+	177	1.75	14.3	Dryer K.O. at 10:04 PM.

(continued)

TABLE F-3B (continued)
DATE: 5/29/79

Time	BED	STEAM			FEED			Comments
	TEMP °F	TEMP °F	PRESS PSIG	MOTOR AMPS	TEMP °F	RATE GPM	PRESS PSI	
12:00 Midnt.	310	362	152	60-68	176	1.5	11.4	
2:00 AM	312	362	150	56-65	176	1.4	11.6	
4:00 AM	310	362	152	58-65	178	1.5	11.5	
6:00 AM	310	362	153	59-65	177	1.5	11.5	
8:00 AM	322	363	156	57-63	177	1.6	11.6	
10:00 AM	322	363	152	57-62	179	1.6	11.6	Dryer shut down for inspection.
12:00 Noon	330	364	152	56-58	168	1.55	5.3	11:00, dryer restored, at 1:00 moisture .02%.
2:00 PM	354	365	154	59-64	168	1.5	5.7	3 sprays.
4:00 PM	321	363	156	65-74	176	1.5	5.7	4 sprays.
6:00 PM	322	363	157	67-74	177	1.5	6.1	5:00 took samples.
8:00 PM	323	363	156	67-74	177	1.6	6.05	
10:00 PM	323	363	156	66-73	176	1.6	6.0	Took sample for moisture dryer steaming.

(continued)

TABLE F-3B (continued)
DATE: 5/30/79

	BED	STEAM			FEED			
Time	TEMP °F	TEMP °F	PRESS PSIG	MOTOR AMPS	TEMP °F	RATE GPM	PRESS PSI	Comments
12:00 Midnt.	310	362	150	65-73	176	1.7	6.3	
2:00 AM	310	362	150	65-72	174	1.7	6.3	
4:00 AM	312	362	152	65-71	175	1.7	6.3	
6:00 AM	312	362	152	65-71	178	1.7	6.3	
8:00 AM	320	362	152	65-71	178	1.7	6.3	
10:00 AM	318	362	152	65-71	180	1.7	7.2	
12:00 Noon	317	362	152	65-71	181	1.7	7.2	
2:00 PM	311	362	152	65-71	180	1.8	8.2	
4:00 PM	308	362	154	65-71	180	1.8	8.3	
6:00 PM	327	362	154	64-70	176	1.5	8.4	Cut back to 3 spaces at 5:00.
8:00 PM	329	363	156	63-68	179	1.5	8.4	
10:00 PM	329	362	155	63-68	180	1.5	8.4	

(continued)

TABLE F-3B (continued)
DATE: 5/31/79

Time	BED	STEAM			FEED			Comments
	TEMP °F	TEMP °F	PRESS PSIG	MOTOR AMPS	TEMP °F	RATE GPM	PRESS PSI	
12:00 Midnt.	329	362	154	61-68	180	1.5	8.4	
2:00 AM	329	362	154	62-68	178	1.5	8.0	
4:00 AM	330	363	155	62-68	177	1.5	8.4	
6:00 AM	329	362	155	62-68	176	1.5	8.4	
8:00 AM	328	362	152	62-68	175	1.5	8.4	
10:00 AM	329	362	152	62-68	176	1.5	8.4	
12:00 Noon								
2:00 PM								
4:00 PM								

APPENDIX G

BENEFITS OF INCREASING THE ABSORPTION EFFICIENCY

APPENDIX G

BENEFITS OF INCREASING THE ABSORPTION EFFICIENCY

Summary

This section presents the results determined by calculation of a study which defines the maximum practical limits of the SO_2 removal attainable from modifications of the Wellman-Lord FGD process. Specifically, the economic and technical impact and the impact on energy consumption of certain process modifications and equipment changes aimed at maximizing removal efficiency are estimated.

The Wellman-Lord/Allied Demonstration Plant at Northern Indiana Public Service Company (NIPSCO) is installed on a 115-MW coal-fired boiler. The FGD unit is designed to remove 90% of the SO_2 when burning coal with a nominal sulfur content of 3.15%. The absorber has three absorption stages. Using the NIPSCO configuration as a base condition, three process modifications were considered:

(1) Add one additional absorption stage to obtain higher absorption efficiencies.

(2) Treat only part of the flue gas at the higher absorption efficiencies.

(3) Obtain higher absorption efficiencies by adding the sodium carbonate makeup solution directly onto the top stage.

These process modifications have the potential for either a further

reduction in SO_2 emissions or savings in the cost and energy demands of the Wellman-Lord process.

Effect on Emission Levels--

The impact of adding the fourth absorption stage was considered for coal sulfur contents ranging from 1 to 4.5%. The upper limits of absorption efficiency were found to vary from about 94.3% for the 1% sulfur case to 97.4% for the 4.5% sulfur case. However, the solution feed rates at these levels of removal were excessive and resulted in substantial increases in costs. A solution feed rate of 110% of the three-stage 90% removal case was chosen as the maximum practical limit for SO_2 removal. At these feed rates, the limits of absorption efficiency were found to be 93.1% for the 1% sulfur case to 96.0% for the 4.5% sulfur case. The equivalent SO_2 emission rates varied from 51 ng/J (0.12 lb/10^6 Btu) to 140 ng/J (0.33 lb/10^6 Btu) for the same range of coal sulfur content. For 3.15% sulfur in the coal, the SO_2 emission rate was 110 ng/J (0.25 lb/10^6 Btu) versus 260 ng/J (0.60 lb/10^6 Btu) for 90% SO_2 removal efficiency.

To achieve the benefits of the higher efficiency absorption, more energy must be consumed. For a coal sulfur level of 3.15% the energy requirements increased 17% chiefly as a result of increased heat input for SO_2 recovery and reduction of the SO_2 to sulfur. Annualized costs for four-stage operation were 4% higher than for three-stage operation.

Effect of Partial Flue Gas Treatment--

The second objective of this investigation was to determine the advantages to be gained from high-efficiency absorption when utilized in conjunction with partial treatment of the flue gas. With this operating mode, energy consumptions and costs will be reduced or moderated through reduced throughputs and reduced reheat energy requirements. To reduce the reheat requirement, advantage is taken of a temperature of the bypassed flue gas about 76.6°C (170°F) higher than the absorber outlet temperature. The situation where only enough flue gas was treated to meet the New Source Performance Standard (NSPS) of 520 ng/J (1.2 lb $SO_2/10^6$ Btu) was considered.

The removal efficiencies required to meet the NSPS are less than 90% for all coal sulfur levels up to and above 4.5%. Therefore, compared to a 90% removal efficiency base case, high-efficiency absorption will only provide part of the benefits if the application of partial treatment is applied just to meet the NSPS. It is apparent that the inlet SO_2 levels will have a profound effect on the amount of flue gas that must be treated. At a 90% SO_2 removal efficiency, the amount of flue gas that must be treated varies asymptotically from 36% for a 1% coal sulfur level to 95% for the 4.5% coal sulfur level. The amount of flue gas which must be treated for a 3.15% sulfur level is 88% and is reduced to 83% with four-stage absorption.

High-efficiency absorption with a fourth stage will also provide an amount of reheat for a recombined gas stream that is comparable to the amount of flue gas that can be bypassed. For an absorber outlet temperature of 52.8°C (127°F), the combined gas stream can be increased to 69.4°C (157°F) (16.6°C (30°F) of reheat) for the 3.15% coal sulfur level case. If it is assumed that the exhaust temperature must be 82.2°C (180°F) to provide adequate plume dispersion and to eliminate the visible plume, the amount of reheat energy required is only 43% of the reheat energy required for full flue gas treatment. No reheat is required up to a coal sulfur level of about 2%.

Energy consumed in the FGD plant with partial flue gas treatment is 76% of the energy consumed with four-stage full-treatment operation. It is 88% of the energy consumed with three-stage operation at a 90% SO_2 removal efficiency (with 3.15% sulfur in coal). Boiler derating varies within a range of 12 to 14% for the three modes of operation.

The reductions in overall costs attributable to partial treatment are not substantial for coal sulfur levels from 3.15% to 4.5%. At the 3.15% coal sulfur level, annualized costs are 92% of costs for four-stage full-treatment operation and 96% of costs for three-stage operation. However, at the 1% coal sulfur level with substantially less of the flue gas to be treated, the reduction in annualized cost is 28% when going from three stages and 90% removal to four stages and partial treatment. Cost estimates were repeated for a boiler size of 500 MW and are discussed in this report.

Effect of Solution Rerouting--

As an approximation, absorption efficiencies will increase from 95.6% for conventional four-stage operation to about 97% for operation with the sodium carbonate makeup fed directly to the fourth (top) stage and with the rest of the feed solution going to the third stage.

Background

This section presents the results of a study which defines for the EPA and potential users of the Wellman-Lord process the maximum practical limits of high-efficiency SO_2 removal. Specifically, the economic and technical impact and the impact on energy consumption of certain process modifications and equipment changes aimed at maximizing removal efficiency are estimated. The analysis is based on the existing Wellman-Lord/Allied Chemical flue gas desulfurization (FGD) facility installed on Unit No. 11 (Mitchell No. 11) at Northern Indiana Public Service Company's D. H. Mitchell Station in Gary, Indiana. The Wellman-Lord SO_2 recovery process employs a sodium base aqueous solution for removal of the SO_2 and thermal regeneration for SO_2 recovery. For conversion of the SO_2, the Allied Chemical SO_2 reduction process for producing elemental sulfur from the recovered SO_2 is used. The results can also be used by the EPA to determine the viability of modifying the existing Wellman-Lord/Allied Installation at NIPSCO for demonstrating the feasibility of high-efficiency SO_2 removal.

FGD units that employ aqueous scrubbing solutions for the removal of SO_2 are typically designed to meet customer requirements resulting from the emission or air quality standards in effect. It is to be expected that some conservatism will be designed into the FGD unit to allow for fluctuations in the sulfur content of the fuel, in boiler load, for temporary losses in SO_2 removal efficiency, or for future tightening of emission requirements. The Federal new source performance standard (NSPS) is used extensively by the states as a limiting standard of compliance for the emission of SO_2 and other pollutants. For coal-fired steam generators, the standard is 520 ng/J (1.2 lb SO_2 per million Btu) heat input. (This was the standard in effect at the time of this study and does not represent current new source standards.) The removal efficiencies required at various levels of coal sulfur content to just meet NSPS are as follows:

Sulfur in Coal, %	Removal Efficiency, %
1.0	34
2.0	67
3.5	81
4.5	85

For many reasons, a utility company may elect to install an FGD unit capable of, for instance, 90% SO_2 removal, even though this system provides them with a considerable margin of safety for the fuel currently available and for the emissions standard currently in effect.

The Wellman-Lord/Allied Chemical FGD unit at NIPSCO is designed to remove 90% of the SO_2 when burning coal with sulfur contents up to 3.5%. An average of about 3.15% is assumed to be the normal concentration of sulfur in the coal. Removal efficiency can be increased by:

- Adding more absorption stages,
- Feeding all makeup sodium carbonate or caustic onto the top absorption stage, and
- Maintaining higher levels of active sodium in the absorbing solution.

This study is concerned primarily with the impact of the addition of one absorption stage. The additional benefit in increased SO_2 removal by feeding makeup sodium carbonate to the added top stage will also be determined. At NIPSCO, the absorber was designed so that an additional absorption stage could be installed in the future for the purpose of demonstrating SO_2 removal efficiencies above 90%.

A disadvantage of FGD processes that use an aqueous scrubbing medium is the visible plume and lack of the bouyancy needed to attain maximum dispersion of the plume. This is a result of the treated flue gas being emitted at or near its saturation temperature. A substantial amount of fuel is needed to raise the flue gas temperature prior to release to the atmosphere. If direct heat is used, the fuel must be natural gas or low sulfur fuel oil to prevent recontamination of the flue gas. In this study, the possibility of utilizing a bypass stream of untreated flue gas for reheat is examined.

Task Purpose--

Three objectives have been established for analyzing the impact of high-efficiency SO_2 removal. First, it is desirable to know the maximum SO_2 removal capability of the Wellman-Lord SO_2 recovery process so that the total reduction of SO_2 and sulfates in the atmosphere from the use of this process can be assessed. As part of this objective, the purpose is to determine the incremental increase in dollars and in energy consumption as the cost for cleanup of the atmosphere by high-efficiency SO_2 removal. The second objective is to assess the feasibility, economic impact, and energy consumption of utilizing high-efficiency SO_2 removal to allow bypassing of a portion of the untreated flue gas without exceeding the emission standard. In this partial treatment scheme, the untreated and treated flue gas are recombined at the exit of the absorber so that the sensible heat of the untreated flue gas is utilized for reheat. The third objective is to provide EPA with the information needed to determine the benefits of installing an additional absorption stage and possible bypass on the NIPSCO unit.

Process Description--

The major steps of the Wellman-Lord Process for flue gas desulfurization are well documented and will not be repeated in great detail here. The following description will focus on the effect on the process of adding additional absorption stages. The total FGD process includes the Allied Chemical process for the reduction of the recovered SO_2 to a salable elemental sulfur product, as configured for the Mitchell No. 11 installation, although it is possible to convert the SO_2 to sulfuric acid or liquid SO_2 in other process configurations.

The major elements of the process are shown in Figure G-1.

At NIPSCO, the flue gas first enters a prescrubber to remove additional particulate matter not collected by the boiler's electrostatic precipitator. It then enters the bottom of the absorber and is contacted with the sodium base scrubbing solution flowing countercurrent to the gas across three mass transfer stages. The primary reaction involved in SO_2 absorption is as follows:

$$Na_2SO_3 + SO_2 + H_2O \longrightarrow 2NaHSO_3$$

The treated flue gas is exhausted to the stack at saturation conditions at the absorber temperature. Without stack gas reheat, the bouyancy effects are limited for obtaining adequate dispersion of the plume. However, stack gas reheat may be made available.

The SO_2-rich absorbing solution leaves the bottom of the absorber and is sent to a single-effect evaporator. The solution is heated in the evaporator and the absorption reaction is reversed, driving off SO_2 and water. After condensing out the water vapor, the SO_2 gas is sent to the Allied Chemical unit where it is reduced to elemental sulfur by a proprietary process that uses natural gas as a reductant.

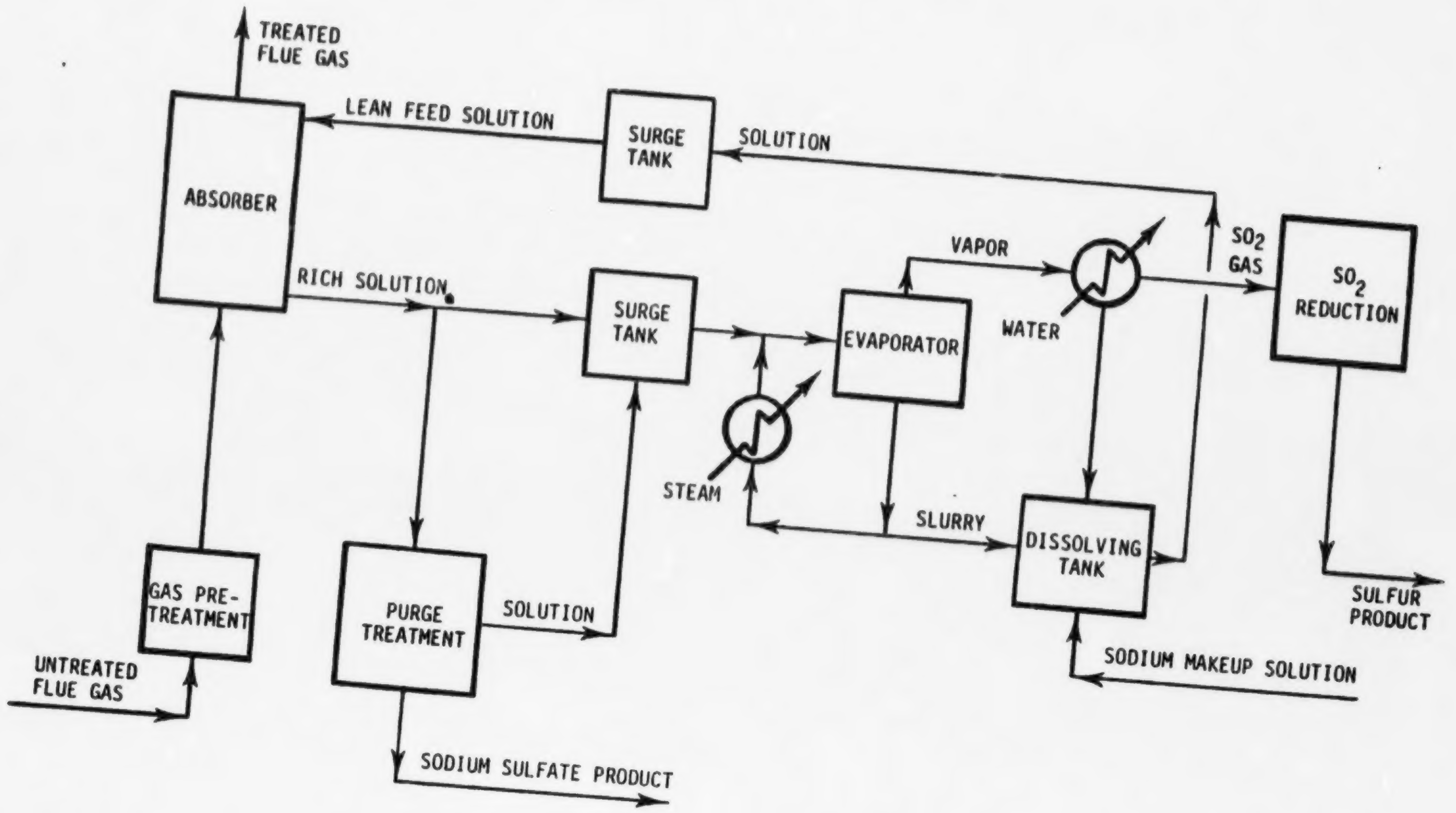

TREATED FLUE GAS
LEAN FEED SOLUTION
SURGE TANK
SOLUTION
ABSORBER
RICH SOLUTION
SURGE TANK
VAPOR
EVAPORATOR
WATER
SO_2 GAS
SO_2 REDUCTION
STEAM
SLURRY
DISSOLVING TANK
GAS PRE-TREATMENT
PURGE TREATMENT
SOLUTION
UNTREATED FLUE GAS
SODIUM SULFATE PRODUCT
SODIUM MAKEUP SOLUTION
SULFUR PRODUCT

Figure G-1. Wellman-Lord/Allied FGD Process.

During absorption, a small portion of the sodium sulfite is oxidized to sodium sulfate; thus there must be a purge from the absorbing solution loop to keep the nonactive sodium at a minimum. This purge stream is centrifuged and dried to produce a sodium sulfate product. The sodium values lost from the absorbing solution are replaced by adding makeup sodium carbonate.

The evaporator bottoms are mixed with makeup sodium carbonate and returned to the top stage of the absorber. The solution is recirculated across each absorption stage; thus solution feed rates can be varied within limits to control the amount of SO_2 removed. By adding a fourth stage, removal efficiencies can be increased at the same solution feed rate and can be further increased by increasing the solution rate. Any increase in removal efficiency will increase the load on the SO_2 recovery and reduction areas. The load on the purge treatment area is a function of the oxidation rate. Higher removal efficiencies are also achieved by maintaining a higher level of active sodium in the system. However, for this investigation, the feed solution active sodium was not varied.

To remove additional SO_2, it is possible to add the makeup sodium carbonate to the fourth (top) stage and the circulating feed solution to the third stage. Advantage can then be taken of the high-absorption activity of the sodium carbonate on a gas stream dilute in SO_2, which, as a consequence of its low SO_2 concentration, has a low-driving force for mass transfer.

The incremental cost of operation is determined from the following major raw material and utilities requirements:

- Makeup sodium carbonate
- Process natural gas
- Cooling water
- Steam for evaporation and for purge drying
- Reheat fuel
- Electricity for blower, pump and miscellaneous drives

For this investigation, electrical drives are assumed for estimating energy consumption and costs, although, at NIPSCO, steam turbines were used on the booster blower, evaporator circulating pump, and SO_2 compressor. Installations with all electrical drives are expected to be more typical of future designs. However, steam can be used if assurance is given for a dependable supply.

Presentation of Results

The objectives of the study are restated as follows:

1. Determine impact on emissions of an additional absorption stage. Specifically,
   (a) Determine the reduction in SO_2 emissions.
   (b) Estimate the incremental increase in cost for the additional cleanup.
   (c) Estimate the incremental increase in energy consumption for the additional cleanup.
2. Assess the feasibility of using high-efficiency SO_2 removal (four absorption stages) to allow partial treatment of the flue gas and thus utilize the sensible heat of a bypassed flue gas stream for stack gas reheat. Specifically,
   (a) Determine the degrees of reheat possible to just meet NSPS.

(b) Determine limits on amount of bypass to raise stack gas exhaust temperature to 82.2°C (180°F).

(c) Estimate the incremental changes in costs and energy consumption.

3. Provide the information needed to determine the benefits of installing an additional absorption stage and possible bypass on the NIPSCO unit.

Two capacity levels were selected for analysis: the existing NIPSCO unit, which treats gas from a 115 MW boiler burning coal with an average sulfur content of 3.15%, and retrofit on a 500-MW unit. The 500-MW unit is essentially the one configured for a detailed cost comparison of selected FGD process conducted by TVA[16], the important variations from the NIPSCO case being coal sulfur content (3.5%) and particulate removal equipment.

To obtain additional SO_2 removal, makeup sodium carbonate can be fed directly to the fourth stage (top stage) of the absorber rather than being premixed with the absorber feed solution as is now done. The effect of this liquor rerouting scheme was analyzed for the case where only carbonate is fed to the fourth stage and only regenerated absorber feed solution is fed to the third stage.

Determination of Process Variables--

Davy McKee computerized process design equations were utilized to determine the effects of four stage design and of varying the inlet sulfur. For assessing the total impact of SO_2 removal performance, economic effects, and energy consumption, the effect on process variables such as absorber solution feed rates, evaporator overhead, purge rates, sodium carbonate makeup rates, and utilities consumption were determined. Mass transfer during the absorption step was determined from vendor-specified tray efficiencies and was based on vapor-liquid equilibrium data published by Johnstone[17] for the SO_2-sodium sulfite/bisulfite system. Inlet flue gas parameters were those used for design of the NIPSCO unit and are tabulated in Table G-1.

TABLE G-1. OPERATING AND COST FACTORS

		115 MW	500 MW
Boiler Annual Load Factor, %		80	80
Boiler Annual On-Stream Time, h		8064	8064
Boiler Heat Rate, Btu/kWh		9237	9237
Boiler Heat Input, 10^6 Btu/h		849.8	3694.8
Boiler kWh Annual Rate		741.9 x 106	3225.6 x 106
Natural Gas Costs, $/$10^6$ Btu		1.642	1.642
Steam Costs, $/$10^6$ Btu		1.32	1.32
Electricity, $/$10^6$ Btu		6.44	6.44
Flue Gas Composition:			
Sulfur in Coal, %	1	3.15	4.5
N_2, mol %	75.6	75.6	75.6
O_2, mol %	5.2	5.2	5.2
SO_2, mol %	0.0695	0.2185	0.3119
SO_3, mol %	0.0011	0.0035	0.0049
H_2O, mol %	6.8	6.8	6.8
CO_2, mol %	12.1	12.1	12.1

Cost Estimation--

It was not an objective of this study to establish detailed cost estimates for all of the high-efficiency absorption and partial treatment cases studied. However, it was necessary to establish a relative effect on costs of the process changes considered. Cost estimates were made as accurately as possible within the scope of the investigation but rough-order-of-magnitude accuracy limits still prevail.

Costs were estimated for both the 115-MW and 500-MW capacity levels. The estimation method proceeded from the establishment of a base case for 90% SO_2 removal on a 115-MW boiler. Actual and estimated cost data available for the Mitchell No. 11 installation could then be used for extrapolation to four-stage absorption and to other coal sulfur levels. Equipment costs scaling was based on throughput rates and rough-order-of-magnitude estimates were made according to the following formula:

$$\text{Cost Scaling Factor} = \frac{\text{Estimate Case Throughput}^{n}}{\text{Base Case Throughput}}$$

$$n = 0.6 \text{ for scale-up}$$

$$n = 0.5 \text{ for scale-down}$$

The incremental costs of operation are strongly influenced by energy costs. A single fuel, natural gas, was assumed to be the source of all heat energy required for the process although availability considerations might rule out gas as a fuel. Natural gas was chosen because of readily available current cost data and because the calculations were simplified by putting all heat energy costs on a common basis. High-pressure steam from the boiler can replace natural gas for all of the heat energy requirements. A natural gas cost of $\$1.642/10^6$ Btu was used.

The ground rules used by TVA were also used to estimate the indirect costs applied to the capitalization and operating costs. Cost indices from Chemical Engineering were used to escalate the costs to February 1977 dollars. Details of the cost estimation methods are included in Exhibit 1 of this appendix.

Impact on Emissions

Three inlet sulfur levels were selected for the analysis. These were coal sulfur contents of 1%, 3.15%, and 4.5%. The median sulfur level (3.15%) reflects the typical sulfur content of coal burned in Mitchell No. 11 and the basis for design of the FGD demonstration unit installed on that boiler. The upper limits of performance found for four-stage absorbers fell within the ranges shown in Table G-2. For the 3.15% coal sulfur level, emission rates of 69 to 110 ng/J (0.16 to 0.25 lb $SO_2/10^6$ Btu) are achievable.

TABLE G-2. MAXIMUM SO_2 REMOVAL FOR FOUR-STAGE ABSORBERS

% in Coal	% Removal	SO_2 Emissions, ng/J (lb/10^6 Btu)
1	93.1 to 94.3	52-43 (0.12 to 0.10)
3.15	95.6 to 97.2	110-69 (0.25 to 0.16)
4.5	96.0 to 97.4	140-90 (0.33 to 0.21)

The Wellman-Lord three-stage absorber installed on Mitchell No. 11 is designed for a removal efficiency of 90%, which is equivalent to an emission rate of 260 ng/J (0.60 lb $SO_2/10^6$ Btu) for a coal sulfur content of 3.15%. Thus, a further reduction in emissions of 60% or more may be achieved with a fourth stage. For cost and energy consumption comparisons, 90% removal with three stages has been used as the base condition for making the comparisons.

Removal performance has been presented as a range of values in Table G-2 because it is difficult to define precisely a practical upper limit for SO_2 removal efficiency. Figure G-2 shows removal efficiency as a function of the absorber solution feed rate. It is apparent that there is a limiting solution feed rate above which there will be little additional SO_2 removed. The lower limit of the removal range for four-stage absorption was selected as the removal obtained with the solution feed rate 10% higher than for 90% removal with three stages. The upper limit was selected as points along the flattened part of the curve of Figure G-2. The annual operating costs increase substantially if, to improve SO_2 removal, solution feed rates are increased above 110% of the base feed rates. This is shown from the relative cost data of Table G-3.

TABLE G-3. COST EFFECTS WITHIN MAXIMUM SO_2 REMOVAL RANGE

	% of Base		
Design/Operating Conditions	Feed Solution	SO_2 Emitted	Annual Costs
Three Stage, 90% Removal (Base Condition)	100	100	100
Four Stage, 95.6% Removal	110	42	104
Four Stage, 97.2% Removal	159	27	123

*115-MW boiler, 3.15% sulfur in coal.

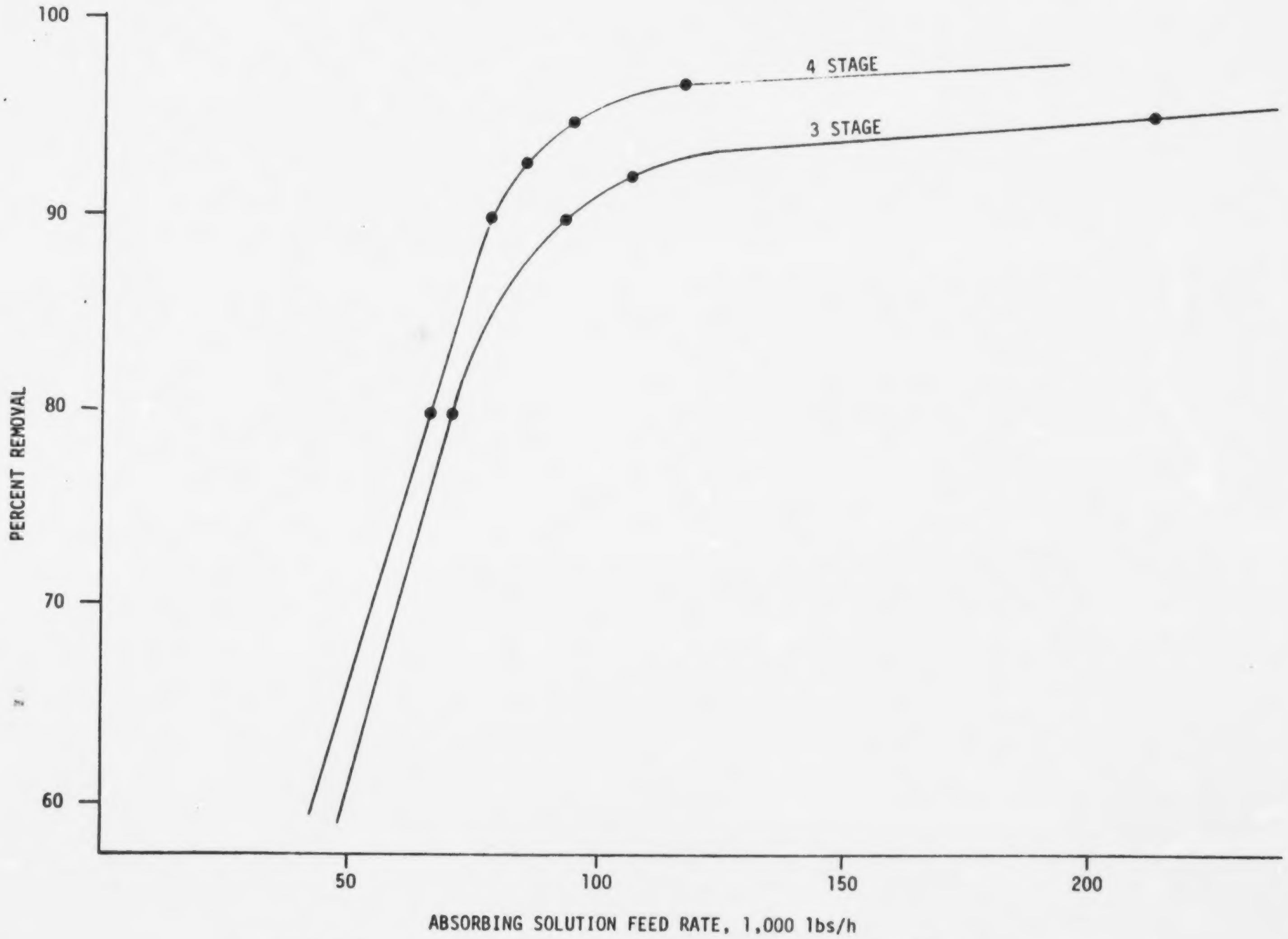


Figure G-2. Effect of liquor flow rate on removal efficiency.

Partial Treatment

High-efficiency SO_2 removal can be utilized to minimize SO_2 emissions as described in the preceding section. Where emission standards are easily met or exceeded, it may be advantageous to utilize the added SO_2 removal performance to allow the treatment of only part of the flue gas so that a recombined stream of treated and untreated flue gas will continue to meet the applicable standards. The advantages are:

- Reductions in cost from reduced absorber and gas handling equipment sizes, and
- Reductions in cost and energy consumption by augmenting stack gas reheat with the sensible heat of the bypassed flue gas.

In this section, we will examine the beneficial effects on costs, reheat requirements, and total energy requirements resulting from partial flue gas treatment when enhanced by high-efficiency absorption. Partial treatment has been limited to treatment that results in emissions not in excess of the current NSPS emission limit, which is 520 ng/J (1.2 lb $SO_2/10^6$ Btu). The reheat energy requirement is that needed to increase the stack temperature from 52.8°C (127°F) and the absorber outlet temperature to 82.2°C (180°F). The inlet flue gas temperature is assumed to be 148.9°C (300°F).

Partial Treatment Limits--

The amount of flue gas that must be treated to meet NSPS will obviously depend upon the inlet SO_2 levels. For the base case (three-stage absorption and 90% removal efficiency), the amount of treatment varies from none for a coal sulfur level of about 0.7% to 95% of the flue gas for a coal sulfur level of 4.5% (Figure G-3). More than 50 percent of the flue gas must be treated for any sulfur levels over about 1.2%. At the 3.15% level of coal sulfur, 88% of the flue gas must be treated. To meet NSPS and simultaneously raise the temperature of the recombined flue gas fo 82.2°C (180°F), 69% of a flue gas from burning coal of about 1.8% sulfur content must be treated. The 69% treatment level applies to any inlet sulfur level if the emission rate is allowed to vary.

With the increased removal efficiencies attained with four-stage absorption, the sulfur content of the coal can only be increased to about 2.0% while still meeting NSPS and still raising the stack gas exhaust to 82.2°C (180°F) (Figure G-4). This indicates that the addition of a fourth stage is not very effective for meeting these conditions.

It is desirable, then, to know: (1) what exhaust temperatures can be achieved while still meeting NSPS and (2) what amount of stack reheat is needed to augment the reheat obtained by partial treatment in order to achieve an exhaust temperature of 82.2°C (180°F) and still meet NSPS. Table G-4 displays the limits imposed by the NSPS emission limitation:

- On the amount of untreated gas which can be bypassed,
- On the stack gas temperature which can be reached without reheat, and
- On the energy required for additional reheat to attain an 82.2°C (180°F) stack gas exhaust temperature.

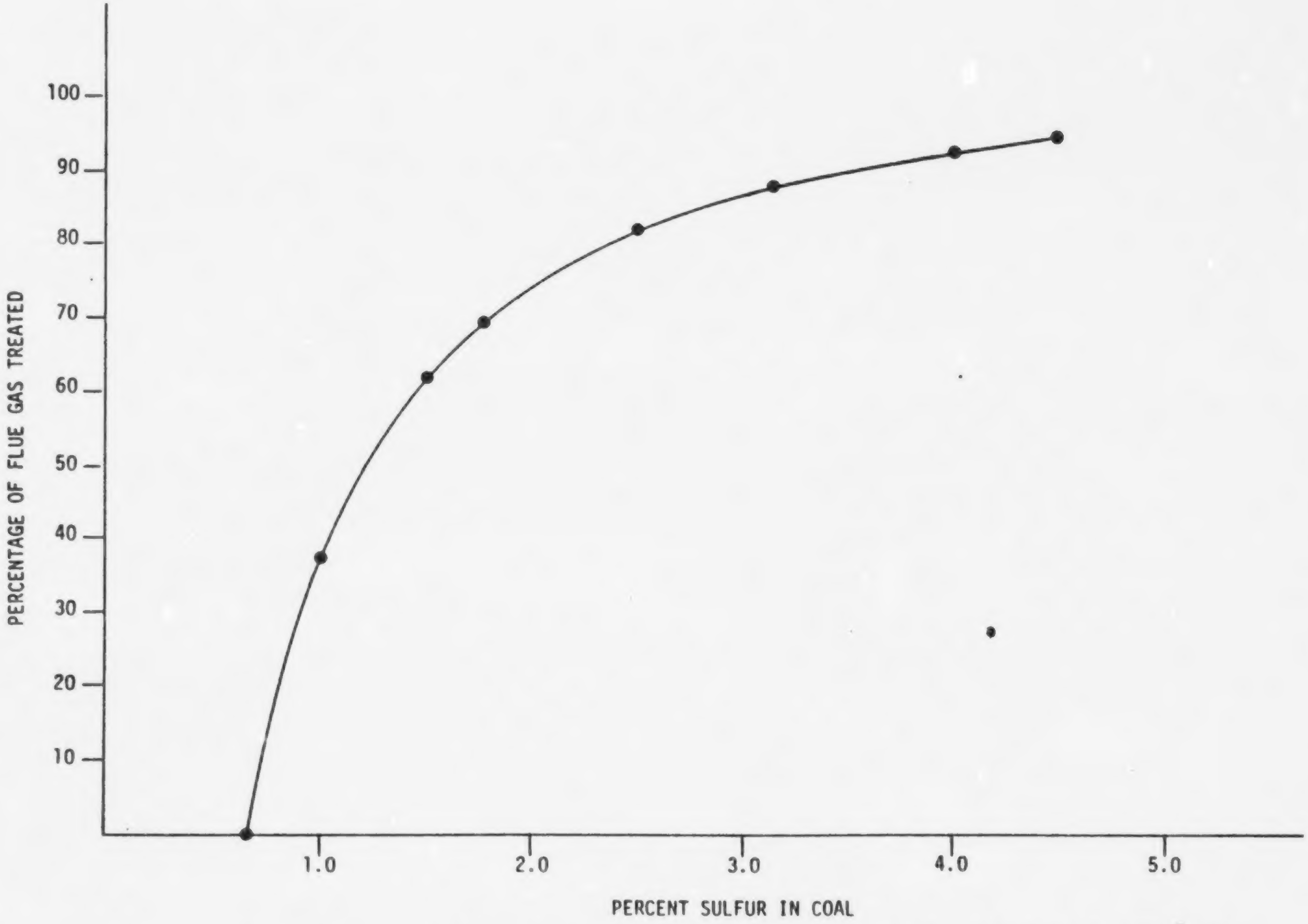


Figure 8-3. Limits of absorber bypass to meet NSPS of 1.2 lb $SO_2/10^6$ BTU for a 90% SO_2 removal efficiency.

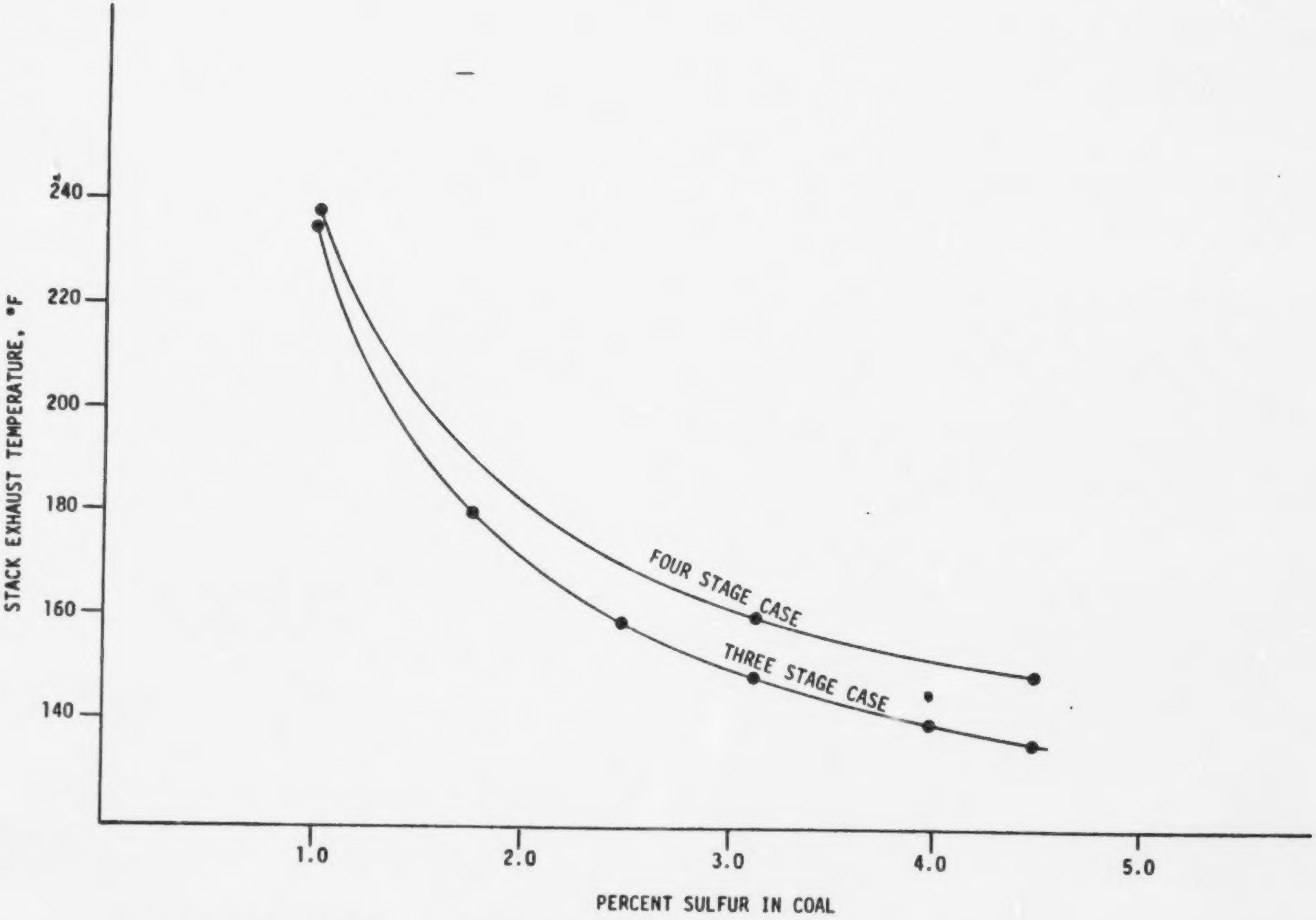


Figure G-4. Effect on stack exhaust temperature from partial treatment without reheat for an emission rate of 1.2 lb $SO_2/10^6$ Btu.

TABLE G-4. PARTIAL TREATMENT LIMITS FOR 115 MW BOILER*

Operating Condition	% Sulfur In Coal	Removal Efficiency, %	Flue Gas Treated, %	Stack Exhaust ^{o}C, (^{o}F)	Reheat Energy To Attain 82.2^{o}C (180^{o}F) 10^6 J/s 10^6 Btu/h)
Three Stage Base Case	3.15	90	88	65.0 (149)	2.26 (7.70)
Four Stage Maximum Removal	3.15	95.6 to 97.2	83 to 81	69.4 - 71.1 (157 - 160)	1.67 - 1.46 (5.71 - 4.97)
Three Stage Base Case	1	90	37	112.8 (235)	None
Four Stage Maximum Removal	1	93.1 to 94.3	36	114.4 (238)	None
Three Stage Base Case	1.8	90	69	82.2 (180)	None
Three Stage Base Case	4.5	90	95	57.8 (136)	3.201 (10.93)
Four Stage Maximum Removal	4.5	96.0 to 97.4	88 to 89	63.9 - 65.0 (147 - 149)	2.40 - 2.26 (8.19 - 7.70)

Note: Reheat energy required with no bypass and absorber exhaust heated from 127°F to 180°F is 3.854 J/s (13.16 x 10^6 Btu/h)

*Limits are based on meeting an emission standard of 520 ng/J (1.2 lb $SO_2/10^6$) Btu by the combined gas stream.

The effect of inlet sulfur and the improvements from four-stage absorption on stack gas exhaust temperatures are shown graphically in Figure G-4. The reduction in the amount of treatment required compared to the three-stage base case varies from 1% for a coal sulfur level of 1 to 7% for a coal sulfur level of 4.5%. The amount of gas treated to achieve a stack temperature of 82.2°C (180°F) is 69%. With four-stage operation, the coal sulfur can be increased from 1.8% to slightly above 2%. As the sulfur level increases, the ability to increase the inlet sulfur level by going from three- to four-stage operation is enhanced. For example, it can be estimated from Figure G-4 that at 71.1°C (160°F) exhaust temperature coal with 3.15% sulfur can be used with four absorption stages whereas a coal sulfur of only about 2.4% is possible with three absorption stages. For sulfur levels of 3.15% and 4.5%, the maximum stack exhaust temperatures increase 21-25% of the maximum increase desired from 52.8°C-82.2°C (127°F to 180°F) when going from three-stage to four-stage operation. Assuming indirect heating and negligible heat losses, the reheat energy requirements are virtually linear with temperature when the temperatures are within the relatively narrow range from 52.8°C to 82.2°C (127°F to 180°F). Table G-5 summarizes the advantages of using partial treatment and high-efficiency absorption to reduce stack reheat requirements.

Effect of Partial Treatment on Energy Consumption--

Partial treatment will reduce energy requirements as a result of reductions in throughput and reheat. However, this is to offset to a degree the energy needed for high efficiency absorption. The additional

TABLE G-5. SUMMARY OF PARTIAL TREATMENT EFFECTS ON EXHAUST TEMPERATURE AND STACK REHEAT REQUIREMENTS[a]

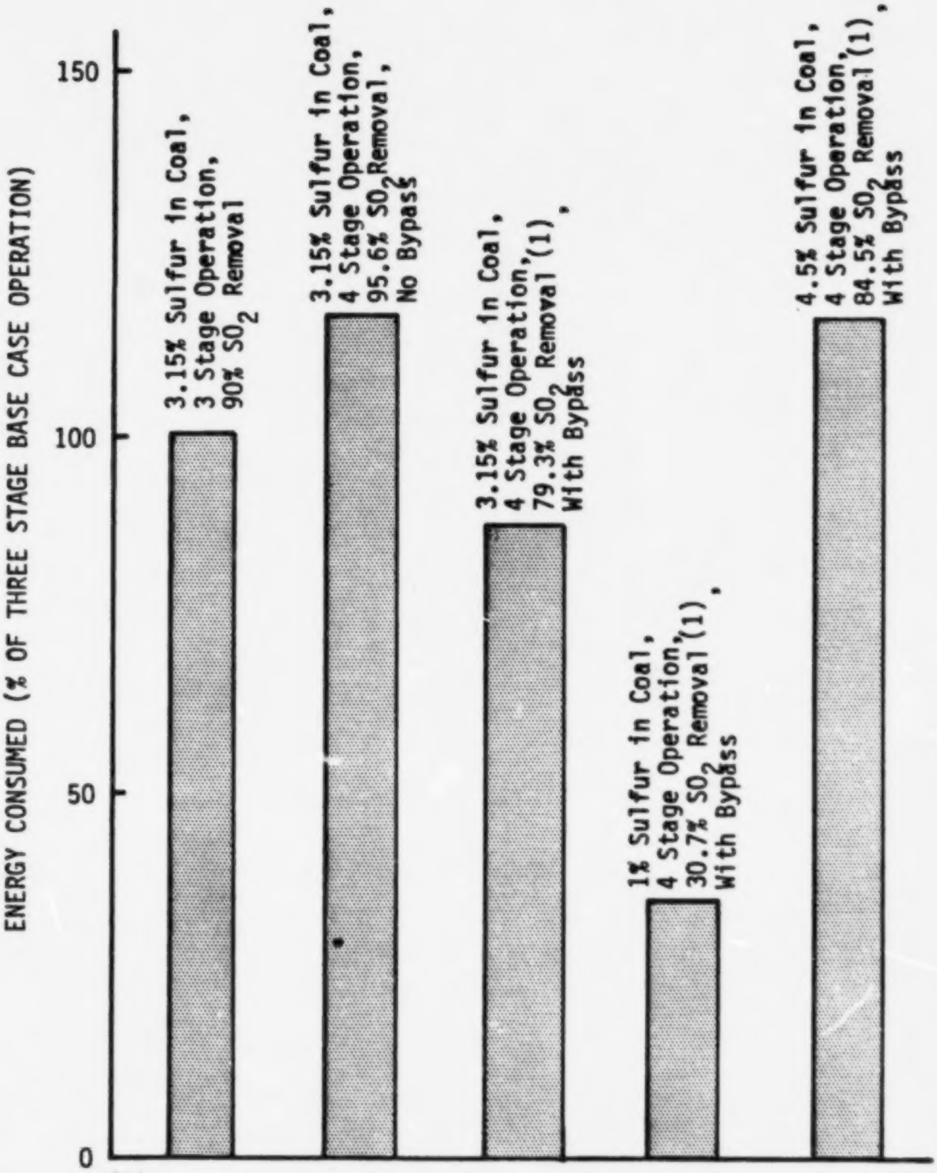
Operating Condition	% Sulfur In Coal	Temperature Increase Above 52.8°C (127°F)		Reheat Energy Required, %[b]
Three Stage Base Case	3.15	12.2	(22)	59
Four Stage Maximum Removal	3.15	18.3	(33)	38
Three Stage Base Case	1	60.0	(108)	None
Four Stage Maximum Removal	1	61.7	(111)	None
Three Stage Base Case	4.5	5.0	(9)	83
Four Stage Maximum Removal	4.5	12.2	(22)	59

[a] For compliance with an NSPS of 520 ng/J (1.2 lb $SO_2/10^6$ Btu).

[b] Reheat energy required =

$$\frac{\text{Reheat energy to makeup to } 82.2^{\circ}\text{C } (180^{\circ}\text{F})}{\text{Reheat energy required with no flue gas bypass}} \times 100$$

energy needed over the requirements for three-stage operation are primarily for the increased booster blower discharge pressure (to overcome the additional pressure drop across the fourth stage) and for the increased evaporation requirements. If four-stage operation is employed, the increase in feed solution rate is limited to 10%, and only enough gas is treated to meet NSPS, the energy consumed is 12% less than that consumed for the three-stage base case (Figure G-5). However, without partial treatment, energy consumption is 17% higher than the



(1) **Based on Total Flue Gas.**

Figure G-5. Effects of high efficiency absorption and partial flue gas treatment on energy consumption.

energy consumed with three-stage operation. The benefit gained from consuming 17% more energy is the reduction in emissions rate from 258 to 107 ng/J (0.60 lb $SO_2/10^6$ Btu to 0.25 lb $SO_2/10^6$ Btu).

The energy consumption effects reported above are for coal sulfur levels of 3.15%. The sulfur level of the coal has a profound effect on energy consumption. For example, for the same four-stage operation and partial treatment case described above, concentrations of SO_2 from burning 1% sulfur coal require only 35% as much energy as three-stage operation with 3.15% sulfur coal. On the other hand, coal with 4.5% sulfur requires 16% more energy to attain the same emissions rate. The effect of inlet sulfur concentrations on energy consumption for the same mode of operation is compared in Table G-6.

TABLE G-6. EFFECT OF SULFUR CONTENT ON ENERGY CONSUMPTION FOR FOUR STAGE OPERATION AND A PLANT EMPLOYING PARTIAL TREATMENT[a]

% Sulfur In Coal	10^9 J/year	Energy Consumed (10^6 Btu/year)[b]	
1	373	(354)	(40%)
3.15	937	(889)	(100%)
3.15	1064	(1,009)[c]	(113%)
4.5	1230	(1,167)	(131%)

[a] Enough gas treated to meet NSPS of 520 ng/J (1.2 lb $SO_2/10^6$ Btu). Solution feed rate limited to 10% over feed rate for three stage operation and 90% SO_2 removal.

[b] For a 115-MW boiler.

[c] For three-stage operation.

Assuming that all of the fuel (except process natural gas) used in the process and all of the electrical energy consumed represents energy not available for power distribution, a boiler derating value can be determined. These values are shown in Table G-7. Again, the high sensitivity of energy requirements to coal sulfur levels is shown.

TABLE G-7. BOILER DERATING EFFECTS FROM HIGH EFFICIENCY ABSORPTION AND PARTIAL FLUE GAS TREATMENT[a]

Operating Condition	% Sulfur In Coal	Energy Consumed[b] / Boiler Heat Input Btu/Btu
Three Stage Base Case, 90% SO_2 Removal	3.15	0.13
Four Stages, 95.6% SO_2 Removal Without Bypass	3.15	0.14
Four Stages, 79.3% SO_2 Removal With Bypass[c]	3.15	0.12
Four Stages, 30.7% SO_2 Removal With Bypass[c]	1	0.05
Four Stages, 84.5% Removal With Bypass[c]	4.5	0.15

[a] Enough gas treated to meet NSPS of 520 ng/J (1.2 lb $SO_2/10^6$ Btu). Solution feed rate limited to 10% over feed rate for three stage operation and 90% SO_2 removal.

[b] Does not include process natural gas.

[c] Based on total flue gas.

Effect of Partial Treatment on Costs--

Both the capital and operating costs of additional stack gas cleanup by adding a fourth absorption stage will necessarily be greater than these costs for three-stage operation. Capital costs will increase due to purchase and installation costs of the fourth stage and equipment cost increases associated with scaleup of the SO_2 recovery and reduction areas to process the additional SO_2 removed from the flue gas. Increased operating costs are the result of increased raw material, cooling water, and energy usage in the recovery and reduction areas. When partial treatment along with four-stage desorption is employed, the flue gas throughput decreases. Most of the process equipment in the absorber area can be scaled down. However, the booster blower would require higher discharge pressure with the same throughput. In the SO_2 recovery area, a slight increase in capacity might be required to accommodate the increase in solution circulation rate if this is required. Virtually all of the process equipment can be scaled down. However, in the SO_2 recovery and reduction areas, some of the capacity must be added back to accommodate the increases in recovered SO_2.

It is estimated that the annualized operating cost for going from three-stage operation to four-stage operation with full flue gas treatment will increase only about 4% for a 115-MW boiler. The estimates of fixed charges are based on the capitalization costs for the demonstration unit installed on Mitchell No. 11 and reported by NIPSCO in March 1976.[6] Operating labor costs are based on Davy Powergas estimates[13] and raw material, cooling water, and energy costs are determined from Davy's proprietary process design data. The appropriate scaling factors and cost indices have been used to adjust these costs for throughput variations and for adjusting all costs to February 1977 dollars. Details of the methods used are given in Exhibit 1.

The costs of partial treatment with four-stage absorption are slightly lower than with full flue gas treatment with three stage operation at the 3.15% sulfur level. In fact, the costs become progressively lower down to the 1% sulfur level (Figure G-6). For a coal sulfur level of 3.15%, the reduction in costs is 4%. Compared to a base cost for three-stage operation and 90% SO_2 removal, the relative costs are shown in Table G-8.

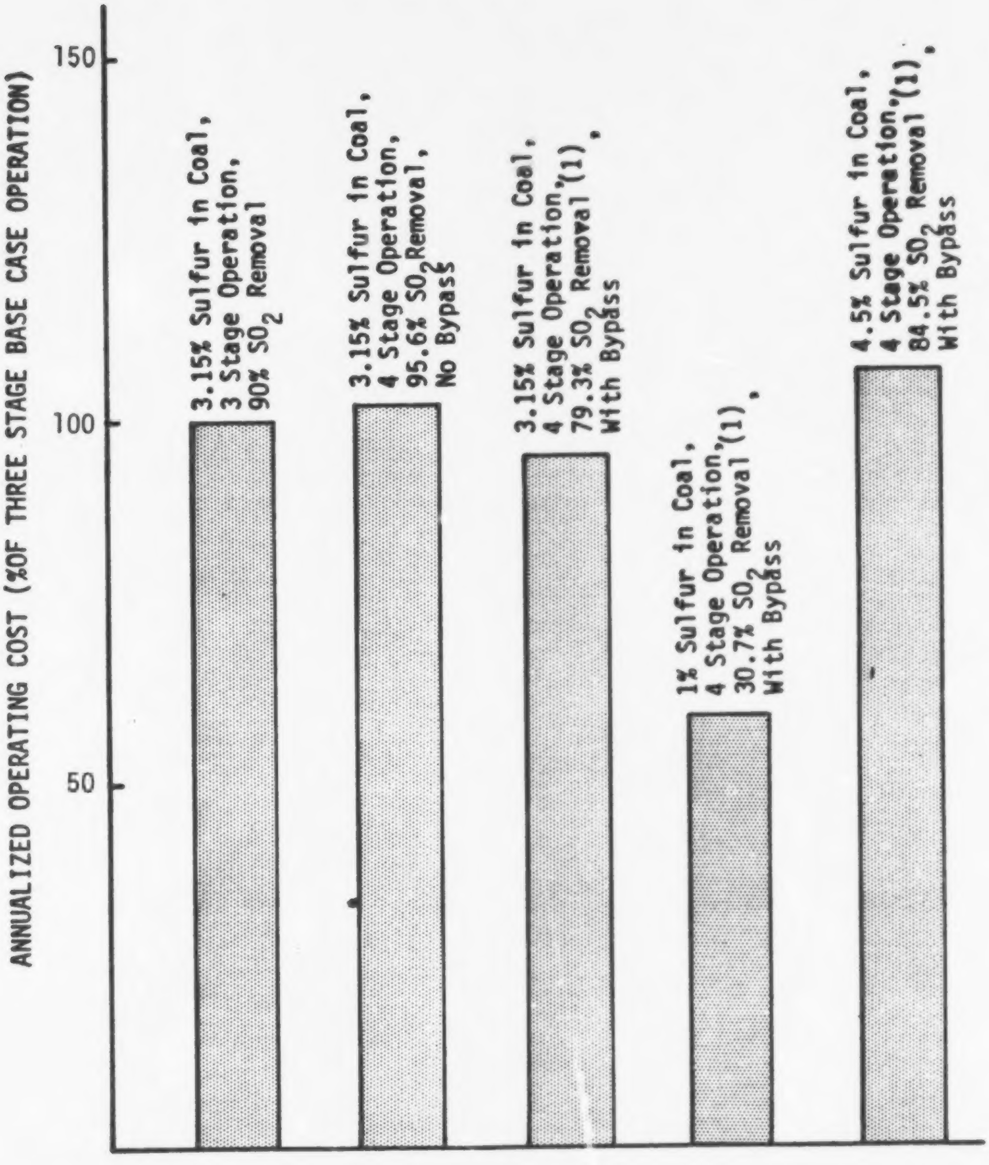
TABLE G-8. COMPARATIVE COST FACTORS FOR VARIOUS TREATMENT MODES (115-MW BOILER, 3.15% SULFUR IN COAL)

Operating Condition	Cost Factor
Three Stages, 90% SO_2 Removal	1.00
Four Stages, 95.6% SO_2 Removal No Flue Gas Bypass	1.04
Four Stages, 79.3% SO_2 Removal,* With Flue Gas Bypass	0.96

*Based on total flue gas.

Relative Costs of 115-MW and 500-MW Installations--

Table G-9 summarizes the cost estimates for all cases studied. Included in the table are cost estimates for high-efficiency absorption and partial treatment employed on a 500-MW boiler. Annualized costs for the 500-MW cases compared to the 115-MW cases are lower, reflecting a smaller fixed investment per megawatt. Raw material and utility



(1) See Figure G-5

Figure G-6. Effects of high efficiency absorption and partial flue gas treatment on annualized operating costs.

TABLE G-9. ANNUAL OPERATING COSTS FOR FOUR STAGE ABSORPTION AND PARTIAL FLUE GAS TREATMENT

	% Sulfur In Coal	Energy Cost MM $	Total Net Operating Cost, MM $	Annualized Cost MM $	Annualized Cost Mill/KWH
115 MW, 3 Stage Base Case	3.15	2.06	2.76	5.70	7.7
115 MW, 4 Stages Without Flue Gas Bypass	3.15	2.18	2.88	5.94	8.0
115 MW, 4 Stages With Flue Gas Bypass	3.15	1.80	2.56	5.45	7.4
115 MW, 3 Stage Base Case	1	1.38	2.23	4.68	6.3
115 MW, 4 Stages With Flue Gas Bypass	1	0.73	1.48	3.38	4.6
115 MW, 3 Stage Base Case	4.5	2.58	3.17	6.48	8.7
115 MW, 4 Stages With Flue Gas Bypass	4.5	2.33	2.93	6.21	8.4
500 MW, 3 Stage Base Case	3.15	9.10	9.99	19.33	6.0
500 MW, 4 Stages Without Flue Gas Bypass	3.15	11.71	12.57	22.93	7.1
500 MW, 4 Stages With Flue Gas Bypass	3.15	7.98	8.91	18.02	5.6

(Continued)

TABLE G-9 (Continued)

	% Sulfur In Coal	Energy Cost MM $	Total Net Operating Cost, MM $	Annualized Cost	
				MM $	Mill/KWH
500 MW, 4 Stages With Flue Gas Bypass	1	3.24	4.71	10.68	3.3
500 MW, 4 Stages With Flue Gas Bypass	4.5	10.29	10.61	20.84	6.5

MM: Million

consumptions and product credits were scaled up from the 115-MW cases to the 500-MW cases in direct proportion to the amount of flue gas treated. On-line times of 8,064 hours at 80% of full load was assumed for both boiler sizes, resulting in annual load factors of 73.6%.

Relative Costs of Three-Stage and Four-Stage Operation--

The changes in annualized costs in going from three-stage operation with 90% SO_2 removal to four stages with partial treatment and increased removal efficiencies are shown in Table G-10.

TABLE G-10. RELATIVE COSTS OF THREE-STAGE AND FOUR-STAGE OPERATION

Operating Condition	Sulfur In Coal, %	Flue Gas Treated, %	Annualized Cost, Percentage of Three Stage Costs
Four Stages with Partial Treatment, 93.1% SO_2 Removal	1	36	72
Four Stages with Partial Treatment, 95.6% SO_2 Removal	3.15	83	96
Four Stages with Partial Treatment, 96.0% SO_2 Removal	4.5	89	96

At the 1% coal sulfur level, the costs have decreased substantially because of the smaller flue gas throughput and savings in reheat but this has been offset by the higher relative recovery and reduction costs as a result of the higher removal efficiency. At the 3.15% and 4.5% coal sulfur levels, lower throughput rates and reheat requirements have produced some savings but the higher recovery and reduction costs have partially offset these savings.

Solution Rerouting

The absorber feed solution consists of a mixture of concentrated solution from the evaporator condensate, and makeup sodium carbonate. With a fourth absorption stage, the sodium carbonate can be fed to the fourth (top) and the remainder of the feed would go to the third stage. The sodium carbonate will more effectively absorb the low concentrations of SO_2 leaving the third stage than would a sulfite/ bisulfite feed solution.

Using Johnstone's equilibrium data, the approximate SO_2 removal efficiency was calculated to be 97% with solution rerouting versus 95.6% SO_2 removal with conventional four-stage operation. The composition of the combined feeds going to the third and fourth stages was not changed from the composition used for the 90% removal base case with a 3.15% coal sulfur level. However, the solution feed rate was increased by 10%. The effects on emission rates and partial treatment are summarized in Table G-11.

TABLE G-11. EFFECT OF SOLUTION REROUTING

				Partial Treatment		
Operating Condition	Removal Efficiency %	Emission Rate ng SO_2/J	(lb $SO_2/10^6$ Btu)	Flue Gas Treated, %	Stack Exhaust °C (°F)	Reheat Energy Required %
Carbonate Feed to Fourth Stage	97	77	(0.18)	82	90.0 (158)	42
Four Stage Conventional Feed	95.6	110	(0.25)	83	69.4 (157)	43
Three Stage Base Case	90	260	(0.60)	88	65.0 (149)	59

Note: 115-MW boiler, 3.15% sulfur in coal. Four-stage operation with 110% solution feed rates.

EXHIBIT 1

APPENDIX G

EXHIBIT 1

APPENDIX G

COST ESTIMATING PREMISES

In order to facilitate computation of the annualized operating costs for both the 115 MW case and the 500 MW case with varying fuel sulfur contents, it was necessary to make certain assumptions as to how the specific capital cost and operating cost constituents varied with fuel sulfur content.

For simplicity, a base case was determined for both the 115 MW case and the 500 MW case; however, the inputs for each of the base cases were different. In the 115 MW case, the costing data were taken from documentation specific to the NIPSCO plant while, in the 500 MW case, economic parameters were obtained from the TVA study.[16]

For the 115 MW case, the base case was taken to be a coal-fired plant burning 3.15% sulfur coal with 90% sulfur dioxide removal.

The 500 MW case utilized the same operating conditions except that the coal firing rate and ensuing emissions were increased to reflect the increased power output.

The capital investment necessary for both the 115 MW case and the 500 MW case was determined by obtaining costing information for each of the following categories:

1. Soda ash receiving and storage
2. Particulate scrubbers and inlet ducts
3. Sulfur dioxide scrubbers and ducts
4. Stack gas reheat
5. Fans
6. Purge treatment
7. Sulfur dioxide regeneration
8. Sulfur dioxide reduction unit
9. Sulfur storage
10. Utilities
11. Service facilities
12. Construction facilities

However, in the NIPSCO case, the costing data available were not subdivided into the various areas which make up the Wellman-Lord FGD process; thus, it was necessary to obtain this information by extrapolation of data found in the TVA study. This was done by plotting the percent of direct investment each one of the above categories represents versus the power rating for an existing coal-fired plant burning 3.5% sulfur (which is not exactly the same as the NIPSCO case of 3.15% sulfur but the error introduced is considered to be significantly small with respect to the overall system error) and employing controls for 90% sulfur dioxide removal. After having obtained the individual percentages for each constituent of the capital investment, the cost was determined by multiplying the derived percentage by the total capitalization charge as given in Reference (2).

Capital costs for the 500 MW case were taken from the TVA study.[16] Due to the TVA study using a different fuel sulfur content than that used by NIPSCO (3.5% sulfur versus 3.15% sulfur) it was required that the TVA cost estimates be made to reflect the lesser sulfur content in the fuel. To accomplish this, those components affected by the fuel sulfur content were varied using the following formula:

$$\text{Cost for 3.15\% S case} = \left(\frac{3.15}{3.50}\right)^{0.5} \times \text{(cost for 3.5\% S case)}$$

In both cases, it was necessary to update the individual, component costs comprising the direct investment to February, 1977 dollars to reflect present-day costs. The updating procedure was accomplished by utilizing appropriate price indices found in Chemical Engineering. Since each component of the direct investment included both material costs and the cost of labor for construction, it was necessary to subdivide each component of the direct investment into a material constituent and a labor constituent. For the NIPSCO case, this was done by employing documentation particular to NIPSCO which gives the breakdown between material and labor costs and factoring this data up to the updated capitalization charges derived from Reference (17). Since the TVA study already included this breakdown as part of its report, it was only necessary to update the costs to February, 1977 dollars. To update material costs, the price index identified in Chemical Engineering as equipment, machinery, supports was used while for labor costs, the construction labor price index was utilized.

In order to include engineering design and supervision costs, construction field expenses, contractor fees, and contingency costs in the capital costs, the TVA study was again consulted to develop an estimate for these components. For the 115 MW case, it was necessary to derive the percentage of direct investment that these four factors represented. This was done by plotting the percentage of direct investment the four components contribute versus the power rating of an existing coal-fired plant burning 3.5% sulfur coal with 90% sulfur dioxide removal and extrapolating back to the 115 MW case. The result was a 48% add-on to the direct investment and this was assumed to be non-variant with respect to fuel sulfur content.

For the 500 MW case, the TVA study reported that engineering design and supervision, construction field expense, contractor fees, and contingency expenses were 43% of the direct investment.

It should also be noted that the TVA study did not include the cost of particulate scrubbing in their assessment of the capital and operating costs for an existing coal-fired power unit. Thus, since the cost of particulate control was to ascertain the cost by utilizing data applicable to a new coal-fired power unit using 3.5% sulfur with 90% sulfur dioxide control. Because the scrubbing system should not be affected by fuel sulfur content but only by fuel ash content and quantity of coal fired per unit time, it was felt that the cost for the scrubber would be approximated by the new plant cost (this is not strictly true since retrofit labor costs are greater than for a new facility; however, the error is considered small with respect to the total system error).

Once a base case was established for both the 115 MW case and the 500 MW case, the cost of systems employing variations created by a change in the fuel sulfur content were computed. However, certain assumptions had to be made in order to estimate the variations in capital charges due to a varying fuel sulfur content. A major assumption was that the component costs which make up the direct investment were a function of only one variable and these costs are given as follows:

1. Soda ash receiving - function of quantity of sodium purged from system.
2. Particulate scrubber - function of ash content of fuel.
3. Sulfur dioxide scrubber - function of amount of flue gas introduced to absorption tower.
4. Stack gas reheat - function of heat requirement to reheat flue gas to 82.2°C (180°F).
5. Fans - same as sulfur dioxide scrubber.
6. Purge treatment - function of total purge stream removed.
7. Sulfur dioxide regeneration - function of overheat load from evaporator.
8. Sulfur dioxide reduction - function of quantity of sulfur dioxide from the SO_2 compressor.
9. Sulfur storage - same as sulfur dioxide reduction.
10. Utilities - non-variant.
11. Service facilities - non-variant.
12. Construction facilities - non-variant.

The means of escalating each of the individual components of the annualized operating costs to February, 1977 dollars is as follows:

1. Soda ash cost - the wholesale prices, industrial chemical price index was used to inflate the cost as given in Reference (1); the cost was assumed to reflect conditions as of January, 1976.

2. Catalyst - the wholesale prices, industrial chemical price index was also used to update the catalyst cost to February, 1977; the base price of the catalyst was derived from the TVA study.

3. Operating labor and supervision - this category was updated to February, 1977 dollars by calculating the cost increase as a function consisting of 50% of the construction labor price index and 50% of the engineering and supervision price index; the labor rate was taken from the TVA study.

4. Natural gas - as in the case of fuel oil, the same FEA document was used which gave monthly costing information for natural gas in the east north central region from November, 1975 to November, 1976. Again, a regression line was drawn through the data points and extrapolated to February, 1977.

5. Heat credit - because it was assumed that steam costs should be linearly related to the bus bar price of electricity (since any steam consumed in the FGD process cannot be used to produce any electric power), three power companies were contacted and an average cost per kilowatt-hour of internally consumed power was derived. The increased cost of heat credit was then determined by factoring the cost given in the TVA study by the ratio of the present-day cost per kilowatt-hour divided by the cost per kilowatt-hour given in the TVA study.

6. Electricity - the February, 1977 cost of kilowatt-hour was determined by averaging the bus bar costs per kilowatt-hour

as given by three power companies contacted for their production costs.

7. Maintenance - labor and material costs for maintenance were determined by the TVA study to be a certain percentage of direct investment as determined by power output. For the NIPSCO case, this percentage was determined by extrapolation of data given for a plant burning 3.5% sulfur with 90% removal for varying load factors.
8. Analyses - the cost of analyses in February, 1977 dollars were determined by updating NIPSCO data (115 MW case) or figures given in the TVA study (500 MW case) utilizing the engineering and supervision price index (used as an indicator of the increase in professional salaries).
9. Administrative and marketing overhead - this component of the annualized operating cost was updated to February, 1977 dollars by use of the engineering and supervision price index (again a reflection of increases in professional salaries): the base costs were taken from NIPSCO data (115 MW) of the TVA study (500 MW).

VARIATION OF UTILITIES AND DELIVERED RAW MATERIALS WITH FUEL SULFUR CONTENT AND AMOUNT OF BYPASS FOR 115 MW CASE

Since a major task of this study is to determine the effect of utilizing a fourth stage with bypass and reheat to just meet the New Source Performance Standard (NSPS) for sulfur dioxide emissions of 520 ng/J (1.2 lb/10^6 Btu) of heat input for varying fuel sulfur loadings, it was necessary to make certain assumptions in order to efficiently ascertain the effect on certain parameters by altering the fuel sulfur content.

The energy consumption of the various pumps and fans present in the system was assumed to be directly proportional to the throughput; i.e., if the throughput of a pump or fan increases by a factor of two, the energy consumption of that pump or fan increases by a factor of two.

Fuel natural gas consumption was dependent upon two factors - steam generation and flue gas reheat requirements. The requirement for steam production was determined from Davy design requirements while stack gas reheat needs were calculated based upon the amount of flue gas which was diverted to the absorber to meet emissions requirements as determined by the NSPS.

Process natural gas utilization was a function of the quantity of sulfur dioxide absorbed and subsequently recovered by the FGD system since methane is needed as a reductant in the reduction unit. Thus, the natural gas requirements are calculated based upon the sulfur content of the fuel and the amount of flue gas diverted to the absorber as dictated by the NSPS.

Heat credit was determined by the amount of steam present in the process since all condensate is returned to the steam generation system thus lessening the energy demand by the amount of heat present in the condensate.

Process water needs were dependent upon purge makeup needs as well as evaporative losses suffered in the absorption tower and were determined from Davy design data.

Electrical power demands were determined by summation of the consumption rates of the various pumps and fans utilized in the FGD system (for consistency with the TVA study, all pumps and fans were assumed to be electrically-driven). Then, in order to allow for any ancillary equipment as well as everyday electrical requirements for lighting, working environment control, and the like, the power requirements for the fans and pumps were increased by ten percent.

VARIATION OF UTILITIES AND DELIVERED RAW MATERIALS WITH FUEL SULFUR CONTENT AND AMOUNT OF BYPASS FOR 500 MW CASE

To determine the variations of utility, soda ash, and catalyst consumption rates with fuel sulfur content, it was assumed that the changes in the various operating parameters followed the same pathways for the 500 MW case as those in the 115 MW case.

The other operating cost inputs for the 500 MW case were handled in a similar manner by adjusting the percentage change obtained for the 115 MW case to the corresponding 500 MW case.

APPENDIX H

PERFORMANCE DATA COLLECTED BY NIPSCO

(NOVEMBER 1979-JULY 1980)

SECTION H-1

BACKGROUND AND SCOPE

With EPA support for seven and one-half months, NIPSCO continued to operate the FGD plant following completion of the test program in October 1979. The performance data reported in this appendix (Table H-3.1 thru Table H-3.7(d)) were collected by NIPSCO with the help of Allied Corporation (formerly Allied Chemical Corporation), the FGD plant operator. There were only 16 days of good Mode 2 operation during the period extending from November 1979 through July 1980. Boiler and FGD plant overhauls were scheduled for the months of November and December. The boiler was operating on January 1, 1980 but FGD overhaul was not completed until January 22, 1980. The reasons for limited utilization of the FGD plant during the ensuing six months will be explained in the following discussion.

TRW has examined the data collected by NIPSCO and has made comparisons if possible with the data collected during the two-year test program. Considerably more extensive and detailed data were collected during the two-year test program. Without the same monitoring system, NIPSCO was not able to provide data corresponding to the data collected by TRW. The actual differences are indicated in the following discussion. In Section H-2, Examination of Post Test Results, comparisons are made by TRW of the NIPSCO post test results with the two-year test program results. Following this discussion, the NIPSCO data is included in Section H-3, Post Test Results.

SECTION H-2

EXAMINATION OF POST TEST RESULTS

SULFUR DIOXIDE REMOVAL

The removal efficiencies reported (see Table H-3.1) averaged 88%, 24-hour average, for 16 days of operation between April 12, 1980 and May 2, 1980. There was one interruption of nearly four days during the period. Thirty-day averages during the second year of the test program varied from 88 to 93% (Figure 3-2). The sample size of the post test data is probably not sufficient to make comparisons of frequency distributions of the data; however, there are a significant amount of the post test data below 89% SO_2 removal (9 out of 16 days). Only 16% of the test program data was less than 89% SO_2 removal. SO_2 removal post test data varied from 85 to 89%.

The removal efficiencies were determined from calculated SO_2 feed rates and from a SO_2 emission rate based on the measured exit SO_2 concentration and "certain simplifying assumptions including saturation of the exit gases." Moisture added during humidification at the prescrubber is the major diluent. Exit temperature does not vary by much and the exit gas is saturated, allowing a fairly accurate estimate of dilution with water from the published vapor pressure data for water vapor. The post test data indicate an apparent dilution factor of about 1.14, defined as follows:

$$SO_2 \text{ Removal} = 1 - \frac{K \times SO_2 \text{ out}}{SO_2 \text{ in}}$$

$$K = \frac{Q \text{ out}}{Q \text{ in}} = \text{dilution factor}$$

Q = flue gas flow (Qout and Qin at identifical conditions
and (Qout > Qin)

A dilution factor of 1.06 was assumed during the test program when data were not available for more accurate calculations. Use of the smaller dilution factor increases the percent SO_2 removal by about 1%.

Determination of SO_2 feed rate was the major difference in the way that SO_2 removal was determined. During the test program, SO_2 feed rate was determined from SO_2 and O_2 concentrations measured upstream of the tail gas return and flue gas flows calculated from measured coal rates and compositions. Hourly average coal rates were used. The post test SO_2 feed rates were calculated from measured coal rates and a "constant (given)" coal composition. The SO_2 concentration measured downstream of the tail gas return was used to calculate the amount of excess air. Daily coal firing rates were used. The tail gas return adds SO_2, CO_2, moisture and excess air to the flue gas inlet the absorber. The design material balance shows that about 4% of the SO_2 feed to the absorber and only about 0.2% of the oxygen feed to the absorber is from the tail gas. Thus, a fairly good estimate of excess air is obtained by the means employed. The effect of the different methods on SO_2 removal results is hard to assess because there are differences not only in the way SO_2 feed rates are determined but also in the way coal feed rates are measured and in the frequency and location of coal sampling. A significant difference in measured data is the SO_2 concentration in the feed which seems to be significantly higher than would be expected from the addition of the tail gas alone.

FGD PLANT DEPENDABILITY

Operating history of the post test period (November 1979 - July 1980) is quite different from that of the second year of the test program. Following a scheduled overhaul of the boiler, the boiler operating period to be considered extends from January 1 to July 31, 1980 and encompasses 5112 hours of potential FGD operating time. However, the boiler did not operate or was not delivering flue gas or steam within the design limitations of the FGD plant for 41% of those hours. In contrast, the boiler was down or not available for only 6% of the second year of the test program (Figure 3-11). Boiler outage was determined from a tabulation of the data of Table H-3.7(a) of the post test data. Boiler unavailability was determined from the descriptive notes of Figure H-3.7(b) and the times indicated, augmented by the information in Table H-3.2(c). Table H-1 is a compilation of this information.

The FGD plant operated 748 hours. For comparison with the second demonstration year, it is assumed these are Mode 2 hours (full operation), although NIPSCO indicates that some Mode 3 (bypass damper open) operation is included. FGD plant reliability was 25% (hours operated) hours called upon to operate. Overall reliability during the second demonstration year was 61%. Operability was only 20% (hours operated/boiler operating hours) and this reflects the effect of 677 hours of boiler operation outside of FGD design limits during which the FGD plant did not operate. FGD plant utilization was only 15% (hours operated/hours in period) for the 5112 hour operating period. Viability index comparison with the second demonstration year are shown in Table H-2.

ENERGY CONSUMPTION

The following comments are made regarding the post test energy consumption data (Tables H-3.4(a), H-3.4(b)).

(a) See Table H-3.4(a). The Btu equivalent of average hourly steam consumption was 85.292 x 10^6 Btu/h. At design steam conditions (550 psig, 750°F), the average consumption would be about 62,000 lb/h. However, it is believed the steam flowmeter readings were not corrected for temperature and pressure. Average steam consumption for full operation during the second demonstration year was 55,800 lb/h corrected for temperature and pressure.

(b) See Table H-3.4(a). Electrical energy consumption averaged 3.155 x 10^6 Btu/h (924 kwh/h), determined by estimating the evaporator circulating pump requirements and metering the rest of the power consumption. During the second demonstration year, metered energy consumption was 713 kwh/h and 330 kwh/h was used as the estimate for evaporator circulating pump requirements.

(c) See Table H3.4(b). For the four months indicated (February-May), sulfur produced per million Btu equivalent in the natural gas averaged 76.6 lbs/10^6 Btu. This compares unfavorably with the second demonstration year results (137.5 lbs/10^6 Btu) and reflects the poor utilization of the FGD plant.

(d) See Table H-3.4(b). No other relevant comparisons can be made with the second demonstration year from this table.

TABLE H-1. HOURS BOILER DOWN OR UNAVAILABLE: JANUARY-JULY

	Down	Unavailable*
January		
ESP maintenance	68.0	
Turbine CV	5.13	
Lost excitation	3.27	
Condensate quality problem		156
Available: 511.60 hours	76.4	156
February		
Turbine CV	1.05	
Condensate quality problem		150.05
Generator H_2 leak	49.33	
Available: 495.57 hours	50.38	150.05
March		
Generator H_2 leak	18.08	
Turbine stop valve	161.88	
Ground protection relay	7.43	
Available: 556.61 hours	187.39	
April		
Condenser leaks	18.13	
Available: 701.87 hours		
May		
No HP steam		24.17
Low system demand	168.27	
Ground protection relay	2.42	
ESP problem	21.18	
Available: 527.96 hours	191.87	24.17
June		
Low system demand	508.82	
Available: 211.18 hours		
July		
Low system demand	303.87	
Prim. air fan	0.42	
Water in turbine oil tank	93.13	
Burning LS coal		346.58
Available: -0-	397.42	346.58

Total hours available:	3005
Total hours down:	1470
Total hours unavailable:	677

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*Boiler operating outside of FGD plant design limits.

TABLE H-2. VIABILITY INDEX COMPARISONS

	Post Test Period (1/1/80-7/31/80)	Demonstration (10/4/78-10/29/79)
Boiler utilization, %	72	94
FGD reliability, %	25	61
FGD operability, %	20	59
FGD utilization, %	15	56

SODA ASH CONSUMPTION

The data of Table H-3.5(b) seem to indicate that less soda ash was required during the post test period. Table H-3 indicates substantially lower purge rates, during the best operating months of April and May.

TABLE H-3. POST TEST PURGE RATES

Month	Purge rate, mol Na/mol S removed
February	0.223
March	0.181
April	0.108
May	0.102
Second demonstration year, April-October	0.217

Soda ash consumption was probably determined from monthly receipts plus inventory changes and provides only an estimate for individual months. Thus, wide fluctuations in monthly purge rates are likely.

PROCESS ECONOMICS

For the nine months, November 1979 - July 1980, operating and maintenance costs average $431,435/month (Table H-3.5(a)). The monthly average variable cost for the second demonstration year was $354,727, using 1978 costs of raw materials and byproducts. Table H-4 compares the raw material and byproduct costs used. High miscellaneous costs and operation labor during the December-January period have a significant effect on the post test operating costs.

TABLE H-4. RAW MATERIAL AND BYPRODUCT COST COMPARISONS

	Post test	Second demonstration year*
Soda Ash	\$98.10-\$110.00/ton	\$88.04/ton
Sulfur	\$32.50-54.40/LT	\$35.00/LT
Sodium sulfate	None sold	\$45.00/ton

*1978 costs

SECTION H-3

POST TEST RESULTS

Northern Indiana Public Service Company
Corporate Headquarters / 5265 Hohman Avenue / Hammond, Indiana 46320 / Tel 853-5200 (219)

December 23, 1980

Mr. Norman Kaplan
Project Officer
U. S. Environmental Protection Agency
Industrial Environmental Research Laboratory
Research Triangle Park, North Carolina 27711

Dear Mr. Kaplan:

SUBJECT: EPA/NIPSCO Contract 38-02-3154

You requested additional operating data for the SO_2 System and the Host Boiler 11 for the period beginning at the time TRW left the site in the fall of 1979. Following are various categories of data calculated or retrieved from various sources. All SO_2 Plant operating data is based on Mode 2 operation. There were only three hours operation in the period on Mode 3 (bypass open).

Table 3.1	SO_2 Feed, SO_2 Removal, SO_2 Emission, expressed in various manners. This data is based on two good operating periods totaling 16 days from April 12, 1980, - May 1, 1980.
Table 3.2(a)	Viability factors of FGD Plant, November, 1979 - July, 1980.
Table 3.2(b)	FGD Plant Operations Data.
Table 3.2(c)	Causes of FGD "Down Time and Frequency."
Table 3.4(a)	FGD Plant Energy Consumption, April 12, 1980 - May 1, 1980.
Table 3.4(b)	FGD Plant Energy Consumption, January - July, 1980.

Table 3.5(a)	FGD Operation/Maintenance Costs.
Table 3.5(b)	Economics - Soda Ash Consumption.
Table 3.7(a)	Unit 11 Boiler Shutdown Data.
Table 3.7(b)	FGD and Boiler Operating Hours Chart.
Table 3.7(c)	Coal Analysis.
Table 3.7(d)	Flue Gas Temperatures.

The sulfate treatment area has been out of service since July, 1979. Sulfates were piped to the bottom ash area. The discharge from the bottom ash area, of which the SO_2 Plant sulfates are a part, continues to be within the NPDES Permit limitations for that discharge.

The sulfur recovered was sold in prices ranging from $32.50 per ton in January, 1979, to $54.40 per ton in May, 1980.

The soda ash purchased varied from $98.10 in January, 1979, to $110.00 per ton in June, 1980.

A sample from the September, 1980, production was analyzed; the results follow:

Sulfur Analysis

Dry Basis	
Sulfur, %	99.68
Ash, %	0.0015
Carbon, %	0.01
Chlorides, ppm	None Detected
Fluorides, ppm	Less than 0.004
Acidity as H_2SO_4, ppm	Less than or equal to 0.0005
Mercury, ppm	0.08
Heavy metals, as lead, ppm	20.0

There is no reliable data available relating FGD Plant operation to the generator performance since the two rows of blading were lost in the low pressure section of the turbine in October, 1979.

I hope that the enclosed data will be of some value to you.

Very truly yours,

E. L. Mann
E. L. MANN
Executive Manager
Electric Production

ELM:nlc
Enclosures
cc: Messrs. H. P. Lyle
J. W. Dunn
F. W. Link
L. A. Meyer
T. P. Steiner
D. V. Stevens
R. C. Adams - TRW

SIGNIFICANT ASSUMPTIONS

3.1 SO_2 Removal

The inlet conditions to the absorber could not be determined from booster blower RPM and fan curve. Instead, for each day, the coal firing rate to Unit 11 was used to calculate the flue gas generated, assuming ideal gases and a constant (given) coal composition. Certain combustion processes were assumed to be perfect. The results of these calculations were SO_2 feed rate, and a gas composition. Using the measured flue gas SO_2 concentration, the amount of excess air was calculated.

The quantity of SO_2 emission was calculated from the measured exit SO_2 concentration and certain simplifying assumptions including saturation of the exit gases.

The calculated values for SO_2 feed and emission were used to establish the removal rate and efficiency. The other items of 3.1 were obtainable from the above calculations and assumptions.

The time period tabulated, specifically April 12-23 and April 28-May 1, was chosen because it is the only time period in the January-July segment of 1980 for which significant integrated (Mode 2) operation was achieved.

3.4 Energy Consumption

All data was measured except electric consumption by the evaporator drive motor. The latter was estimated from the evaporator operating hours.

3.7 Flue Gas

All data was measured.

3.1 DATA FOR 16 DAYS OF OPERATION DURING THE PERIOD APRIL 12 - MAY 1, 1980

SO_2 REMOVAL	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
SO_2 Feed, lb/hr	5331	4540	4360	5341	4578	4088	6011	4676	3827	4464	4267	4453	6409	4169	5428	4965
SO_2 Emission, lb/hr	672	596	574	707	639	451	665	514	427	495	486	486	749	629	768	669
SO_2 Removal, %	87.4	86.9	86.8	86.8	86.0	89.0	88.9	89.0	88.8	88.9	88.6	89.1	88.3	84.9	85.9	86.5
SO_2 Feed, $>$ 4842 lb/hr	X			X			X						X		X	X
SO_2 Feed, $>$ 5380 lb/hr							X						X			
SO_2 Feed, $<$ 1700 PPM																
SO_2 Removed, lb/hr When FG Flow $<$ 160,000 $>$ 400,000 cfm	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SO_2 Feed, lb/hr When FG Flow $<$ 160,000 $>$ 400,000 cfm	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SO_2 Feed, PPM	2810	2840	2630	2590	2630	2400	2330	2300	2280	2370	2540	2630	2700	2770	2780	2610
SO_2 Emission, PPM	324	340	316	314	332	240	234	230	230	238	262	260	288	382	358	327
SO_2 Emission, lb/10^6 BTU Input To Boiler	0.75	0.78	0.78	0.79	0.83	0.66	0.66	0.66	0.67	0.66	0.68	0.65	0.70	0.90	0.84	0.80
SO_2 Removal, lb/hr	4659	3944	3786	4634	3939	3637	5346	4162	3400	3969	3781	3967	5660	3540	4660	4296

ALL DATA FOR MODE
2 OPERATION,
INDIVIDUAL DAYS

3.2(a) FGD VIABILITY FACTORS

Month	FGD System % Availability	 % Reliability	 % Operability	 % Utilization	Booster Blower % Availability	Absorber % Availability	Evaporator % Availability	SO_2 Reduction Area % Availability
*NOVEMBER, 1979	0	0	0	0	0	0	0	0
*DECEMBER	0	0	0	0	0	0	0	0
*JANUARY, 1980	54	0	0	0	96	93	61	100
FEBRUARY	46	27	27	23	96	100	96	52
MARCH	75	25	20	15	100	100	91	90
APRIL	63	53	52	50	100	100	99	63
MAY	63	26	23	15	100	100	66	97
JUNE	53	0	0	0	73	80	100	100
JULY	100	0	0	0	100	100	100	100

*The FGD System and the host boiler were being overhauled in November and December, 1979. The January, 1980, data is based on 237 hours after completion of the SO_2 Plant overhaul on January 22nd.

3.2(b) FCD PLANT OPERATIONS DATA

	HOURS					
Month	In Period	Available	Called Upon	Operated	Not Operated	Reason Not Operated
November, 1979	720				720	Boiler and FGD Plant overhaul
December, 1979	744				744	Boiler and FGD Plant overhaul
January, 1980	744	129	160	0	504	FGD Plant overhaul completed 22nd
					156	Generating station condensate problem
					48	Start up
					36	Reduction area - sulfur pipe blockage
February, 1980	696	322	611	163	320	Reduction area - sulfur piping blockage Replaced SO_2 South heater Leak in regenerator shell
					150	Generating station condensate problem
					63	Unit 11 off - FGD available
March, 1980	744	561	433	108	319	Reduction Area - Sulfur blockage Incinerator control SO_2 compressor Leak in piping
					179	Unit 11 off - FGD available
					125	Start up
					13	Lost power to FGD plant

3.2(b) FGD PLANT OPERATIONS DATA cont.

	HOURS					
Month	In Period	Available	Called Upon	Operated	Not Operated	Reason Not Operated
April, 1980	720	456	678	362	272	Reduction area - sulfur blockage
					86	Start up
May, 1980	744	467	436	115	243	Boiler not in service - FGD available
					180	Evaporator circulating pump motor
					96	Start up
					86	Reduction area - piping leak SO_2 compressor turbine
					24	High pressure steam not available
June, 1980	720	384	211	0	415	Unit 11 off - FGD available 211 hours
					153	Absorber
					152	Booster blower
July, 1980	744	744	0	0	413	Unit 11 off - low system load
					331	Unit 11 on low sulfur coal

3.2(c) CAUSES OF FGD DOWN TIME & FREQUENCY

Reason For No Operation Of FGD Plant	Hours Jan. 80-July 80 incl.	No. Of Events	% Of Time
Unit 11 off - FGD Available	1202	8	28.0
Reduction Area	1033	12	24.0
FGD Plant Overhaul	504	1	12.0
Start Up	355	17	8.0
Boiler on Low Sulfur Coal - FGD Available	331	1	8.0
Generating Station Condensate Problem	306	2	7.0
Evaporator Circulating Pump Motor	180	1	4.0
Absorber	153	1	4.0
Booster Blower	152	1	4.0
High Pressure Steam Not Available	24	1	0.6
Lost Power to FGD Plant	13	1	0.4
TOTALS	4253	46	100.0

The FGD Plant overhaul, which was started in November, 1979, was completed on January 22, 1980. There are 5112 hours, January, 1980, through the month of July, 1980. The FGD Plant did not operate 4253 hours during that period. There were 46 periods when the FGD Plant did not operate. The reasons for no operation are in 11 classifications.

The classification, "Start Up," is for start ups when the host unit was the reason for the interruption. Time for all start ups resulting from the FGD Plant interruptions are included in the outage time of the reason for no operation.

3.4(a) FGD PLANT ENERGY CONSUMPTION

For 16 Days April 12 - May 1, 1980

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
Steam Cons. BTU/hr x 10^3	82901	87943	84448	86224	87656	82099	82500	84219	84734	83932	82958	86052	88802	89891	91151	79177
Elect. Cons. BTU/hr x 10^3	3413	2986	3271	3413	3271	3271	3413	3129	2560	3271	3413	3271	2986	2844	2986	2986

3.4(b) FGD PLANT ENERGY CONSUMPTION

JANUARY - JULY, 1980

Energy Consumption	Jan.	Feb.	Mar.	Apr.	May	June	July
Electricity, 10^6 BTU/Mo	1,273	2,130	1,755	2,276	1,693	962	843
Steam, 10^6 BTU/Mo*	17,810	48,196	28,762	57,892	24,498	-	-
Natural Gas, 10^6 BTU/Mo	1,450	2,842	2,360	4,536	2,212	1,204	1,277
SO_2 Removed vs. Energy to FGD from Boiler lb $SO_2/10^6$ BTU	-	9.6	9.7	16.9	13.4	-	-
Sulfur Produced vs. Energy from Natural Gas #S/10^6 BTU	0	77.1	56.1	102.6	70.6	0	0

*Emergency Steam Consumption Not Monitored

3.5(a) FGD OPERATION MAINTENANCE COSTS - DOLLARS

	1979		1980						
	Nov.	Dec.	Jan.	Feb.	Mar.	Apr.	May	June	July
NIPSCO Labor - Operation	6099	9637	1839	1943	1169	3062	1434	251	100
NIPSCO Labor - Maintenance	10720	946	2630	16517	789	1083	1511	(17711)	-
Allied & Contract Labor - Operation	73664	107642	110268	73407	108525	100110	94582	91152	103443
Allied & Contract Labor - Maintenance	17792	31973	35651	12571	23024	24810	25468	24425	36014
Chemicals	29900	41525	(30250)	20052	7707	14690	14157	3410	-
Operating Supplies	15214	60289	32854	13203	8144	32720	69511	46558	(2638)
Miscellaneous Expenses	213551	253693	444842	147641	231755	218689	209954	239841	160578
Residual Sales - Sulfur - Credit	-	-	-	(5289)	(4964)	(9422)	(4435)	(1105)	-
Residual Sales - Sodium Salts	-	-	-	-	-	-	-	-	-
Allied Management Fee	30000	31000	31000	29000	31000	30000	31000	30000	31000
TOTAL	396940	536705	628834	309045	407149	415742	443182	416821	328497

AMORTIZATION NOT INCLUDED

3.5(b) ECONOMICS - SODA ASH CONSUMPTION

	January	February	March	April	May	June	July
Soda Ash Consumption vs. Absorber Operation lb Na_2CO_3/hr	501	603	584	511	623	-	-
Soda Ash Consumption vs. SO_2 Removal, lb Na_2CO_3/ lb SO_2	-	0.37	0.30	0.18	0.17	-	-

3.7(a) UNIT 11 BOILER SHUTDOWN DATA

Month	Duration	Reason	Scheduled or Unscheduled
Nov. 79	Entire Month	Unit Overhaul	Scheduled
Dec.	Entire Month	Unit Overhaul and Post Overhaul Tests	Scheduled
Jan. 80	68h - 00m	Precipitator Maintenance	Unscheduled
	5h - 08m	Turbine Control Valve Malfunction	Unscheduled
	3h - 16m	Lost Excitation	Unscheduled
Feb.	1h - 03m	Turbine Control Valve Malfunction	Unscheduled
	49h - 20m	Generator H_2 Leak	Unscheduled
March	18h - 05m	Generator H_2 Leak	Unscheduled
	161h - 53m	Remove Turbine Stop Valve Screens	Scheduled
	7h - 26m	Generator Ground Protection Relay	Unscheduled
April	18h - 08m	Condenser Leaks	Unscheduled
May	2h - 25m	Generator Ground Protection Relay	Unscheduled
	54h - 56m	Dispatcher - Low System Load	Scheduled
	55h - 00m	Dispatcher - Low System Load	Scheduled
	21h - 11m	Precipitator Problem	Unscheduled
	58h - 20m	Dispatcher - Low System Load	Scheduled
June	31h - 18m	Dispatcher - Low System Load	Scheduled
	419h - 20m	Dispatcher - Low System Load	Scheduled

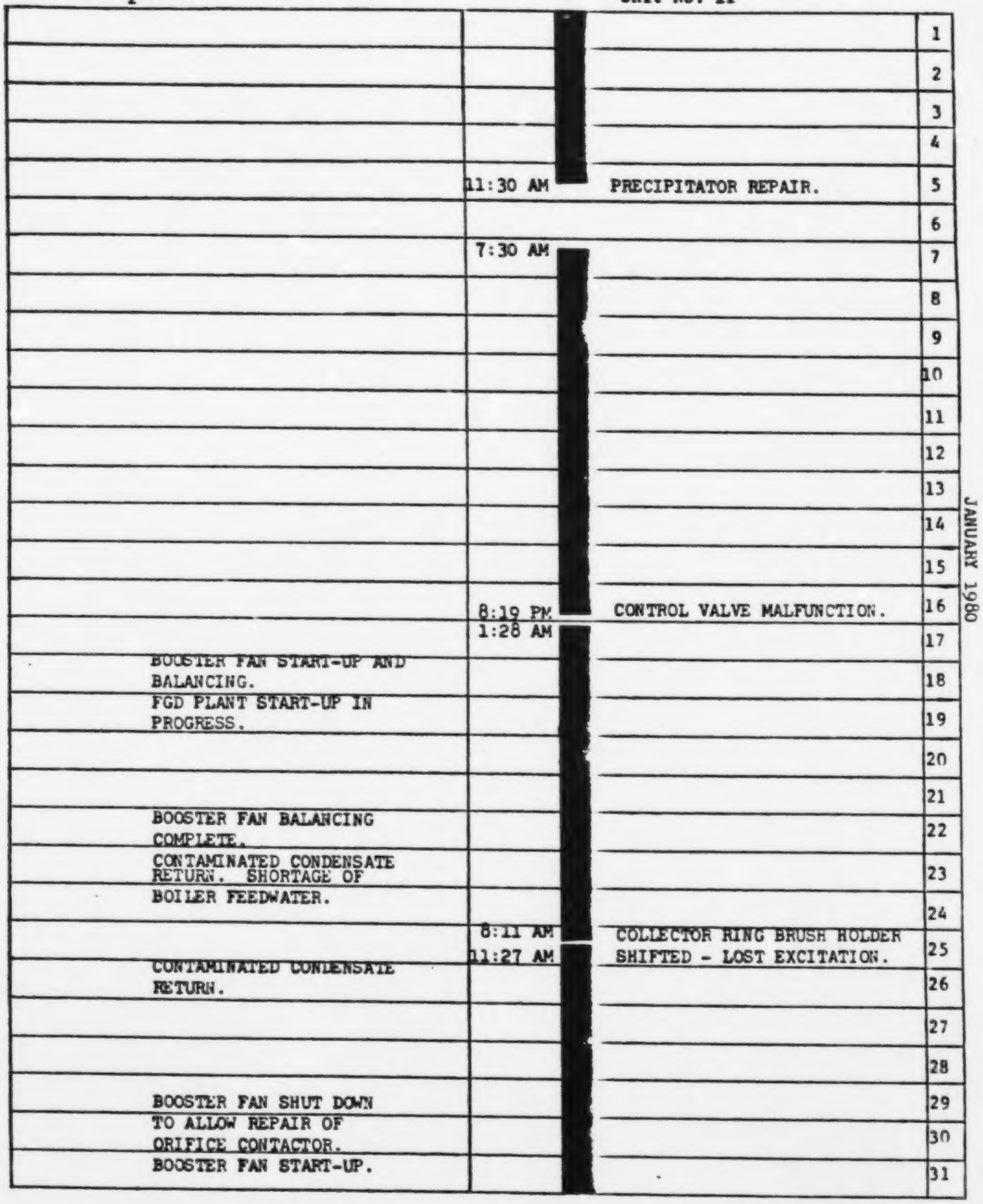
3.7(a) UNIT 11 BOILER SHUTDOWN DATA cont.

Month	Duration	Reason	Scheduled or Unscheduled
June cont.	58h - 11m	Dispatcher - Low System Load	Scheduled
July	177h - 43m	Dispatcher - Low System Load	Scheduled
	0h - 25m	Boiler Primary Air Fan Shorted	Unscheduled
	64h - 17m	Dispatcher - Low System Load	Scheduled
	93h - 08m	Water in Turbine Oil Tank	Unscheduled
	61h - 52m	Dispatcher - Low System Load	Scheduled

3.7(b)

OPERATING HOURS

SO_2 Plant **Unit No. 11**



JANUARY 1980

SO_2 Plant		Unit No. 11	
			1
			2
			3
			4
	11:30 AM	PRECIPITATOR REPAIR.	5
			6
	7:30 AM		7
			8
			9
			10
			11
			12
			13
			14
			15
	8:19 PM	CONTROL VALVE MALFUNCTION.	16
	1:28 AM		17
BOOSTER FAN START-UP AND BALANCING.			18
FGD PLANT START-UP IN PROGRESS.			19
			20
			21
BOOSTER FAN BALANCING COMPLETE.			22
CONTAMINATED CONDENSATE RETURN. SHORTAGE OF BOILER FEEDWATER.			23
			24
	8:11 AM 11:27 AM	COLLECTOR RING BRUSH HOLDER SHIFTED - LOST EXCITATION.	25
CONTAMINATED CONDENSATE RETURN.			26
			27
			28
BOOSTER FAN SHUT DOWN TO ALLOW REPAIR OF ORIFICE CONTACTOR.			29
			30
BOOSTER FAN START-UP.			31

3.7(b)

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

FEBRUARY, 1980			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	Orifice		1
2	Contactor		2
3	Repairs		3
4			4
5			5
6		Off 3:42pm On 4:45pm Turbine Control Valve Malfunction	6
7			7
8	On 11:28am Off 1:45pm		8
9	Piping Leak In		9
10	Reduction Area		10
11	On 1:45pm		11
12	Off 10:25am Leak In Shell Of "A"		12
13	8:00am Regenerator		13
14	SO_2 Plant Available		14
15	But Condensate Quality		15
16	Problem in Generating		16
17	Station		17
18			18
19	On 2:03pm Off 4:15pm		19
20	Leak In Regenerator		20
21	On 9:48am		21
22	Off 8:12am On 10:29am Leak In Regenerator		22
23			23
24			24
25	Off 10:40am On 2:00pm Replace Leaking SO_2 S. Htr.		25
26			26
27	Off 9:15am #11 To Come Off Line	Off 10:40pm Generator H_2 Leak	27
28			28
29			29

3.7(b)

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

MARCH, 1980

DATE	FGD PLANT	UNIT NO. 11	DATE
1	SO_2 Compressor	█	1
2	Problems	█	2
3		█	3
4		█	4
5		█	5
6		█	6
7		█	7
8		█	8
9		█	9
10		█	10
11	On 0:10pm Off 2:20pm █ SO_2 Leak In Reduction	█	11
12	On 0:45pm █ Area Piping	█	12
13	█	█	13
14	█	█	14
15	█	█	15
16	Off 4:22pm █ Incinerator Control	█	16
17	Failure	█	17
18	On 10:20am Off 12:01pm █ Unit 11 Scheduled To	Off 7:09pm █ Remove Turbine Stop	18
19	Be Off	Valve Screens And	19
20		Check H_2 Seals	20
21			21
22			22
23			23
24			24
25		On 1:02pm █	25
26		Off 10:56pm █ False Trip - Ground	26
27		On 6:22am █ Protection Relay	27
28		█	28
29		█	29
30		█	30
31	On 11:45am Off 3:50pm █ Sulfur Blockage In Reduction Area	█	31

3.7(b)

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

APRIL, 1980			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	Sulfur Blockage	▮	1
2	In Reduction Area	▮	2
3	On 11:40am Off 1:40pm ▮ Sulfur Blockage	▮	3
4	In "B" Claus Reactor	▮	4
5		▮	5
6		▮	6
7		▮	7
8		▮	8
9		▮	9
10		▮	10
11		▮	11
12	On 10:15am ▮	▮	12
13	▮	▮	13
14	▮	▮	14
15	▮	▮	15
16	▮	▮	16
17	▮	▮	17
18	▮	▮	18
19	▮	▮	19
20	▮	▮	20
21	▮	▮	21
22	▮	▮	22
23	▮	▮	23
24	Off 1:40pm ▮ Unit 11 Off	Off 3:48pm ▮ Condenser Leaks	24
25		On 9:56am ▮	25
26	SO_2 Plant	▮	26
27	Start Up	▮	27
28	On 11:20am ▮	▮	28
29	▮	▮	29
30	▮	▮	30

3.7(b)

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

MAY, 1980			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	█	█	1
2	Off 1:15pm █	█	2
3	Repair	█	3
4	Bearing	█	4
5	Failure	█	5
6	In The	█	6
7	Evaporator	Off 9:36am On 12:01pm — False Trip - Generator Protection Relay	7
8	Pump Motor	█	8
9	1:15pm — SO_2 Plant Available	█	9
10	1:25pm — But No Steam From Unit 11	█	10
11	SO_2 Plant	█	11
12	Start Up	█	12
13	On 10:30am Off 1:45pm ▬ Reduction Area SO_2	█	13
14	Leak & SO_2 Compressor	█	14
15	Turbine Governor	█	15
16	Steam Leak	█	16
17	— SO_2 Plant	Off 2:32am Dispatcher - Low	17
18	Available	System Load	18
19	— SO_2 Plant Start Up	On 9:28am █	19
20	On 1:05pm █	█	20
21	█	█	21
22	█	█	22
23	█	█	23
24	Off 7:00pm █	█	24
25	SO_2 Plant Available	█	25
26	Low System Load	Off 2:27am Dispatcher - Low	26
27		System Load	27
28		On 9:27am Off 4:29pm ▬ Precipitator Problem	28
29		1:40pm — Available	29
30		Dispatcher - Low	30
31		System Load	31

3.7(b)

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

JUNE, 1980			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	Repair Booster	Dispatcher Low System Load	1
2	Blower Turbine	On 7:18am █	2
3	Governor And	█	3
4	Flush Oil System	█	4
5		█	5
6		█	6
7	— Available If	Off 8:25am █ Orders of Dispatcher	7
8	Needed	Low System Load	8
9			9
10			10
11			11
12			12
13			13
14			14
15			15
16			16
17			17
18			18
19			19
20			20
21			21
22			22
23			23
24	7:45 pm — Repair Absorber	On 7:45pm █	24
25	Tray Spiders And	█	25
26	Caps	█	26
27		█	27
28	1:49 pm — Absorber Repair	Off 1:49pm Dispatcher Low	28
29	Continues	System Load	29
30			30

3.7(b)

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

JULY, 1980			
DATE	FGD PLANT	UNIT NO. 11	DATE
1			1
2			2
3			3
4	No		4
5			5
6	FGD		6
7			7
8	Plant	On 9:43am Off 10:08am On 10:34am █ Pri Air Fan Motor	8
9		█	9
10	Operation	█	10
11		Off 10:30pm █ Dispatcher Low	11
12	in	System Load	12
13			13
14	July.	On 2:47pm █	14
15		█	15
16	Low	█	16
17		█	17
18	Sulfur	Off 9:04am █ Turbine Oil Problem	18
19			19
20	Coal		20
21			21
22	Tests		22
23		On 6:12am █	23
24	Being	█	24
25		Off 4:21pm █ Dispatcher Low	25
26	Made.	System Load	26
27			27
28		On 6:13am █	28
29		█	29
30		█	30
31		█	31

3.7(c) COAL ANALYSIS

Month	Tons	Moisture %	Ash %	Volatile %	Fixed Carbon %	Sulfur %	BTU %
Nov. 79*	0	-	-	-	-	-	-
Dec. 79*	713	13.64	10.34	31.56	44.46	3.01	10,775
Jan. 80	29,110	15.08	12.31	28.93	43.68	2.88	10,269
Feb. 80	26,802	15.64	10.81	28.28	45.27	3.14	10,383
Mar. 80	24,908	15.39	11.16	29.64	43.81	3.05	10,318
Apr. 80	26,229	15.55	9.63	29.71	45.11	2.87	10,392
May 80	22,633	15.52	10.53	29.21	44.73	3.02	10,364
June 80**	0	-	-	-	-	-	-
July 80**	0	-	-	-	-	-	-

Chlorine varies from 0.01 to 0.02% in the as received coal. All data is on "as received" basis. Analyses are averages of samples obtained daily from the coal conveying belt feeding the boiler using an automatic sampler. Sampling was according to ASTM D 2234-72, Type 1, Condition B, Systematic Spacing. The coal was from Illinois Coal Seams 5 and 6, Captain Mines, near Percy, Illinois.

*Unit 11 and SO_2 Plant Overhaul

**Unit 11 on Low Sulfur Coal. SO_2 Plant not needed.

3.7(d) FLUE GAS TEMPERATURES

Data for 16 Days of Operation During the Period April 12 - May 1, 1980

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
Ave. Temp., Orifice Contactor, °F	289	290	289	291	292	292	298	298	287	305	318	299	297	305	301	302
Max. Temp., °F	293	293	294	295	296	300	302	304	292	313	328	305	304	309	305	308
Min. Temp., °F	284	288	287	289	288	278	297	289	281	293	306	285	286	298	298	293